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EVALUATION OF GRANULAR BED DEVICES

Final Report

Contract No. PH-86-67-51

Phase III

AVATD-0107-69-RR

June, 1969

Prepared for

Process Control Engineering Program
National Air Pollution Control Administration
U. S. Public Health Service

Submitted by

Advanced Chemical Processes Section
Avco Applied Technology Division
Lowell, Massachusetts 01851

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ABSTRACT

Recent interest in dry sorption processes for SO_2 removal from hot stack gases has led to investigation of the concept of a dry granular bed device used as a simultaneous SO_2 sorber-flyash collector for power station flue gases. In these devices, an SO_2 sorber such as alkalized alumina would also act as the granular collecting particles for flyash.

At present, it is not possible to accurately predict dust collection performance on granular beds from aerosol filtration theory. Experimental studies of granular filter dust collection have demonstrated that under certain conditions, high efficiencies have been obtained. A survey was undertaken in order to identify the most likely candidates among granular bed devices for power plant services. Most of the devices considered were panel filters, in which granular material filled a narrow vertical shaft. In the crossflow configuration as exemplified by the Dorfan-Impingo Filter or the Carnegie-Mellon shaft filter, the granules steadily fall through the shaft, carrying out with them accumulated dust, while the gas flows through the face of the panel in a crossflow manner. In the Squires designs, the granular material is held in the bed during treatment and intermittently removed from the panel by pneumatic or mechanical means. The Squires GSC design by virtue of appropriate internal spacers affords countercurrent gas-solid contacting. The Fuller and Ducon Companies report having fixed bed granular filters under development. Little detailed data is available on the performance of the above sorber-dust collectors.

Mathematical models for SO_2 sorption on alkalized alumina were developed for counterflow, crossflow, and fixed bed contactors. Calculations using these models indicated that there was no critical problem with regard to removal of SO_2 at conditions which the developer of a particular device considered good for dust collection.

Mechanical layouts were developed for the Squires GSC and crossflow designs to enable preliminary comparative cost estimates to be arrived at. The layouts were oriented toward equipment for an 800 mw coal fired power station.

Cost estimates were made for the devices considered both as sorber-collectors and as dust collectors alone. Values of both capital and operating costs for the granular devices were of the same order as published estimates for other SO_2 removal systems. Estimated costs for granular devices as dust collectors along were of the order of cost for electrostatic precipitation equipment.

Since of the non-proprietary designs, neither the Squires nor crossflow designs have been tested to the extent necessary to demonstrate good mechanical operability, it is recommended that experimental work be undertaken on these two concepts.

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I. INTRODUCTION

A. General Background and Approach

The removal of flyash from flue gas streams has been the subject of great industrial and academic attention. Technical developments and legislation have led to dust collecting devices becoming standard equipment in power stations as well as in many other industrial operations. For high efficiency dust removal from large volumes of hot gas, the devices most commonly in use at present are electrostatic precipitators, baghouse collectors, and wet scrubbers. Collection efficiencies greater than 95% can be achieved with these types of equipment, and there are many different specific designs and configurations available. Frequently, the gas cleaning plant consists of one of the above devices preceded by mechanical cleaning devices such as cyclones whose purpose is to remove the larger particles before the gas reaches the high-efficiency collector.

The patent literature contains descriptions of multitudinous cleaning devices, many relating to variations in electrostatic, bag filter, scrubber, and cyclone designs, and others relating to still different concepts. One of these other concepts is that of a granular bed filter. The majority of the patented ideas in the area of dust collection have probably never been experimentally tested to the extent required to determine their value on an industrial scale, and this certainly applies to the granular filter area.

Broadly speaking, a granular gas filter as considered for the purposes of this report is a stationary device consisting of a bed of separate, relatively close-packed granules which make up the collection surfaces for the filter. In order to prevent the particulate matter collected from plugging the interstices formed between the granules and causing excessive pressure drops, the device should embody some means for either periodic or continuous removal of the flyash from the collecting surfaces. This description, then, excludes fluid or dispersed beds where the granular particles are kept in motion by the gas being treated. It does include fixed bed or close-packed moving bed devices.

Considerable attention has also been focused in recent years on the removal of SO_2 from stack gases from coal-fired power stations through the development of dry processes, operating on hot flue gases. The SO_2 removal can be achieved by sorption of the sulfur species on pellets of several types displaying high specific rates and capacities. Several reactor configurations have been considered and tested such as fixed, moving, dispersed, and fluidized bed devices. Since the dry SO_2 sorption processes involve the use of granular particles for gas separation, the question arose as to whether the same particles could be used as the collecting medium in a granular filter for flyash collection and thus establish devices which could serve as simultaneous sorber-flyash collectors for power stations. This specific concept has been advanced by inventors such as Prof. A. M. Squires of New York, New York and Mr. M. I, Dorfan of Pittsburgh, Pennsylvania with regard to granular bed devices of their own design. In addition, the few granular beds actually considered for commercial sale can also be evaluated for simultaneous sorption-filter service. The particular impetus for this investigation arises from the possibility of a single type of equipment installation being capable of treating a gas stream for both particulate matter and sulfur, and any economic advantages deriving therefrom as compared to separate dust collection and sulfur removal devices.

It was the purpose of this study to evaluate the potential of granular bed devices for both large scale sorption-dust collection and dust collection applications alone from the standpoints of available information on technical performance and of economic potential. The procedures followed during the course of the program were:

1. Survey of available information on granular bed devices
2. Modeling of devices for sorption performance
3. Development of mechanical layouts and configurations
4. Development of preliminary cost estimates

The calculations and estimations were based on a gas cleaning plant for an 800 mw power station. A description of the assumed gas and flyash properties is given in Table I.

Although a number of different sorbents can be considered for SO_2 removal, the discussion in this report will be limited to one of the most extensively studied materials, alkalized alumina. This was done only to establish a specific and realistic basis of discussion since rate and capacity data for this sorbent are available. Other sorbent materials, however, can certainly be used.

B. Experimental Granular Bed Filtration Studies

Before entering into a discussion of the particular granular bed devices treated as part of this study, it is of value to first indicate the status of present aerosol filtration knowledge and then present some discussion of some previous experimental granular bed studies. Although a more detailed treatment of aerosol filtration theory appears in Appendix A, the discussions of filtration theory and granular bed experiments are not intended as complete treatments of their respective topics. They are included for two reasons:

1. No attempt was made in this study to mathematically model the granular devices considered from the standpoint of prediction of dust collection performance. It is desired to indicate why realistic theoretical or extrapolative predictions cannot now be obtained from filtration theory at the present state of development, and

2. On the other hand, several experimental studies of dust

collection in fixed granular beds have been carried out. Since these experiments were not carried out in any of the predominately moving bed devices that will be described in Section II, their interpretation is not obscured by mechanical problems arising from solids motion. It is therefore desired to indicate that these earlier experiments do give the promise of good dust collection performance in granular beds although direct extrapolation to any equipment discussed later in this report would probably not be meaningful.

In brief, the theoretical approach to aerosol filtration is concerned with identification of the controlling mechanism for particle collection on a solid body such as diffusion, direct interception and inertial impurgement, and then, development of expressions for the efficiency of collection. Most of the experimental work has been with regard to fiber filters. From a completely theoretical standpoint, consideration of anything other than particle collection by a single mechanism on a single collecting body is a most difficult problem in the field of fiber filtration, and many investigators have expressed their results in terms of dimensionless correlations. No such general correlations have been developed for granular filters, however. In addition, extrapolation of fiber bed correlations to granules is not feasible because of the size difference between the fibers used (on the order of 10 microns) and granules of industrial interest (6-60 mesh). The difficulty is caused by the fact that the controlling particle dimension for performance of a granular filter may not be the average granule diameter but rather the size of a small protruberance on the surface of the granule.

Among the earliest of granular filter tests were those carried out in 1948-49 at the General Electric Company, Hanford Works (4) (5) (14). Sand bed filters were used to remove radioactive particles from ventilation system air. The total concentration of suspended particles was on the order of 1 gr/1000 ft^3, while the active particles were estimated to be present in concentrations ranging only from 0.01 to 0.0001 gr/1000 ft^3. The particle sizes ranged from 0.5 to 2 microns. The test chambers consisted of vertical cylinders, 10-12 inches in diameter and 3-5 feet long. The filter sand was supported on an 8 mesh screen welded to the sides of the tank upon which was placed successively finer grades of sand. Gas from the main ventilation system was passed through the bed in an upflow manner for the tests. Collection efficiencies were based on the decrease of measured radioactivity of the gas passing through the bed rather than on decrease of total particle concentration.

Correlation of the experimental data obtained as a function of variation in operating conditions was complicated by changes in particle size and size distribution of the active material in the gas from run to run. Particle size changes evidently were capable of overshadowing the effects of other variables such as bed depth and gas velocity.

These tests were run at low velocities, no higher than 10 ft/min., and the controlling mechanisms were most likely diffusion and gravity settling, the latter being another collection mechanism more important at low flow rates. Efficiencies greater than 99% were shown to be possible in these tests and some typical data are summarized in Table II. These data were "admittedly selected to eliminate discordant results and to indicate relatively clean-cut logical trends".(14) The final correlation was

$$N_t = \frac{K L^{\frac{1}{2}}}{V^{1/3} D^{4/3}}$$

Where the efficiency is given by the number of transfer units, (a measure of filter performance similar in concept to the same term used in chemical engineering mass transfer processes)

V = supervicial velocity, fpm

L = sand depth, feet

D = average sand grain diameter, inches

K = coefficient depending on sand type as given in Table III

It was noted that the Hanford sand which gave the highest collection efficiency results displayed the greatest irregularity in shape.

On the basis of these preliminary tests, large scale sand filters were designed and set up at Hanford with a capacity of 30,000 cfm each. The design face velocity used was 6 ft/min., and the filtering layer consisted of a 24 inch depth of 20-40 mesh Hanford sand. Seven layers of coarser sand were also used in the bed to aid in flow distribution. Overall pressure drops were in the range of 4-7 inches of water, and an average collection efficiency for these large units was determined to be about 99.7%. The potential lifetime of these sand beds was estimated at about 5 years based on estimated accumulation of particulates.

Strauss and Thring (7-8) (12) carried out studies of filtration of sub-micron fumes from open hearth furnace gases on granular beds. The units were made up of 2 inch pipe and the granular material, crushed insulating brick, was retained on a stainless steel wire gauze. Tests were carried out on waste gas drawn directly from an operating steel furnace. Here also, the data were obscured by variations in the dust loading and temperature of the gases. These latter factors were dependent on the particular furnace operation being carried out at the time data were taken, and could not be controlled. Typical data are presented in Table IV. Correlation was hampered in that vala of several variables were simultaneously changing from run to run, but efficiencies greater than 90% were obtained at velocities up to about 100 ft/min on relatively large (5/16 in - 7 mesh) collection particles. The controlling mechanism in these tests might be conceived to be inertial impingement, but

Strauss emphasizes the importance of the effect of inlet dust concentration on efficiency(12). It is maintained that dust particle agglomeration in the bed occurring in the tortuous paths between the collecting granules is highly significant to the collection process. On the average, the higher collection efficiencies were found at higher inlet loadings, and this can be observed in Table IV. On the other hand, the magnitude of the outlet concentration was occasionally high also for the higher inlet loadings, e.g., tests No. 8 and No. 10. This effect is not brought into the previously-mentioned dimensionless correlations developed for fiber filtration studies. The fiber beds had porosities of over 90% as compared to the 40% to 60% range that might be expected for a granular filter. Thus, for the same overall volume of filter, the interstices in a fiber bed would be expected to be much larger than for a granular filter and the agglomeration tendency is possibly not as important.

A study of the collection of submicron aerosol particles on electrified granular beds was carried out by Research-Cottrell, Inc., Bound Brook, New Jersey. One such device is described in U. S. Patent No. 2,990,912, July 4, 1961. Collection of particles occurred in a bed of 3-6 mm glass beads held between two screens maintained at different electrical potentials (Figure 1). Upstream of the glass spheres was a zone where the aerosol particles were given an electrical charge. Data taken on collection of a 0.5 - 0.7 micron methylene blue aerosol are shown in Figure 2.(17) It is noted that a 99.5% efficiency was obtained in a bed under two inches in thickness with a pressure drop of 0.20 in wg. and at the relatively high face velocity of 2 ft/sec. Research-Cottrell has not incorporated this concept into a specific filtering device but indicates that it could be compatible with some type of moving bed system. One possibility cited was the maintenance of an electric field across a gas flow duct through which collection granules were falling.

Although the primary collection mechanism in an electrified packed bed is different from the more common ones previously discussed, a description of this study is included as it is another example of the potential of granular filters as high efficiency collectors.

From an overall standpoint, some conclusions can be drawn regarding the potential of granular bed devices as dust collectors. Clearly, insufficient data exist to permit the development of a general correlation that could be used for predicting the performance of any given proposed device. Indeed, owing to the fact that available theory on which such an expression would be based does not take into account interactive phenomena such as flow interference, dust loading on the bed, and agglomeration, it is not particularly desirable at present to strive for such a general correlation.

Experimental difficulties must be recognized as well. Since trends for any correlation are established as the result of a relatively large number of individual runs, it is, of course, required to maintain constant all factors except those to be varied in any given test. Perhaps the one factor most difficult to control from run to run in any realistic industrial dust system is the particle size and distribution in the gas flowing to the filter. This has been a considerable problem in investigations, such as that of Strauss, which have operated on gas streams taken directly from industrial ducting, as compared to those operating with synthetically prepared dust-gas mixtures. A serious difficulty in the latter approach, however, relates to the problem of feeding the dust into the gas stream in a dispersed form. There is no clearcut solution to this dilemma but if an actual process stream is being used to obtain data for a correlation on a given device, it is essential to know the extent of size variations over a period of runs. Again, if the correlation is developed completely on a laboratory basis with dilute monodisperse aerosols as were used for the fiber bed studies, the problem may be avoided if no significant agglomeration occurs; this may not be a practical or economic approach, however, with regard to obtaining filtration data on industrial dusts.

A given device for a range of operating conditions should therefore be tested for its own merit, and the data obtained correlated on a rational basis insofar as possible, but some gross empiricism is probably unavoidable. On the other hand, the limited data which are available for granular filters do indicate that high collection efficiency performance is an attainable goal. Most of the previous work has concentrated on the collection of particles of a far smaller average size than is found in a typical flyash (Table I). Considering that 90% efficiencies have been obtained in a granular bed at relatively high velocities by Strauss with submicron particles, it is perhaps not unreasonable to expect even better performance under the same conditions with larger dust particles. This is based on the assumption that any physical properties of the particles which affect their collection are not drastically different, e.g., electrostatic effects, agglomerative tendencies, etc. (See Section II,C.1)

In summary, the following points can be stated:

1. The technical potential of a granular filter as a flyash collector has not been established.

2. Based on previous work, the concept could have great merit with regard to filtration performance for flyash at industrial gas throughput rates.

3. Reliable values of filtration effeciencies for granular bed devices at present cannot be predicted from aerosol filtration theory.

4. Great care must be taken in filtration experiments to control and evaluate the effects on filter performance of changing particle size and size distribution in the inlet gas.

II. SURVEY OF GRANULAR BED DEVICES

Information on granular beds was obtained from several sources including searches of Chemical Abstracts and the patent literature. Advertising brochures of companies manufacturing gas cleaning equipment were scanned. Many of the leading companies were queried by telephone as to the possible manufacture or development of this type of equipment. Direct contact was also made with likely academic personnel and inventors.

The survey showed that granular bed filters have never come into general use as a preferred mode of industrial gas cleaning. The majority of such devices which have been proposed are found in the patent literature and, with several exceptions, little or no work has been directed toward development of implementation of the patented designs on an industrial scale. Several companies indicated that they had tested proprietary variations of granular bed concepts at one time, such as combination granular-fiber filters or as described in Section I-B, an electrified packed bed, and two other companies do have granular devices under present development.

Beside the work described in Section I, experimental development on specific granular filters has been carried out for the following designs:

1. Dorfan Impingo filter - once commercially sold
2. Consolidation Coal Company crossflow filter
3. Crossflow shaft filter presently under study at Carnegie-Mellon University
4. Squires LS filter (the similar Squires GSC filter has not been experimentally tested but holds promise as a counter-current sorber)
5. The Ducon Company "Expanded Bed" filter, now under development
6. The Fuller Company "Sand Bed" filter, also under present development
7. The Lurgi Company "Gravel Bed" filter once marketed in Europe and undergoing redesign as of late 1968
8. The Granger Filter Company concept, not tested on gas service, but the basis of a commercially available liquid filtration system

The first five designs listed above can be grouped under the heading of shaft or panel filter. By this is meant a narrow vertical enclosure containing the granular material provided with appropriate entry and exit ports to permit the gas to enter, distribute itself throughout the contact zone, and then leave the device. This is shown schematically in Figure 3. The collected dust which must be removed from the granular device to prevent excessive pressure drops is either displaced through the gas entry louvers or is carried downward by the collection granules as they flow out of the bed. This latter distinction can be used to distinguish two categories of panel filters, and these will now be discussed.

A. Crossflow Shaft-Falling Solid Designs

1. Dorfan Impingo Filter

The Dorfan Impingo filter (U. S. Patent 2,604,187) was offered commercially as a dust collector in the 1950's and was described in its old advertising literature as being used, for example, for the collection of asbestos dust at several Canadian installations, for collecting dust, fumes and steam arising from coking operations, and for collecting fumes arising from steel production in an electric furnace. Apparently, it was not tested on power station stack gases.

The device is a vertical panel filter in which the collecting particles continuously fall through the panel. The granule flow rate is controlled by the setting of a rotary valve located near the bottom of the shaft. The falling solids are contacted by a cross-flow stream of gas (Figure 4). Note that the two separate filtering shafts in series are used.

Note that in the Dorfan design the opposite walls of the filtering shaft are not parallel but are slightly offset from the vertical so that the shaft is tapered with its narrowest point at the top and the widest at the bottom. This construction is observable in Figure 4. The collecting particles thus travel downward in a shaft of constantly increasing cross-sectional area. The rationale is that the tapered construction acts to prevent hangups caused by size increases of the collection particles as the dust loading builds upon their surfaces during their contact time with the dirty gas. As the particles move down through the device, the dust collects on the surface of the individual collectors and is carried out of the equipment by their motion. Performance characteristics of the largest unit described are shown in Table V.

The Impingo filter installations are no longer in use and the equipment is not presently marketed. The inventor, M. I. Dorfan of Pittsburgh, makes the claim, however, of 98% collection efficiency for 2-10 micron dust particles in a gas containing 3/8 grains/ft^3 at a face velocity of 6 ft/sec.

2. Consolidation Coal Company Filter

A crossflow filter was studied first on a pilot scale and then with a full size installation by the Consolidation Coal Company in the period 1950-52. This equipment was used for collecting dust from hot gas from coal drying operations. Lump coal, 3/8 to 1-1/4 inches in size, was the collection medium. The design concept for the full size installation is shown in Figure 5.

The collecting solids were fed from an overhead bin to two separate shafts. The shafts at each side consist of six separate cells facing the inlet gas which flowed through the beds from a central duct and exited via ducts at either side of the installation. Typical operating characteristics are given in Table VI.

A very serious operating problem resided in the fact that the coal collecting medium displayed a high attrition rate during passage through the system, necessitating a makeup rate about 8 times the design value. Most of the attrition appeared to be occurring in the vertical conveyor to the feed bin. An attempt to improve this situation was made by switching to a gravel filter medium, but the degradation problem was not greatly improved. By taking into account the coal degradation, however, efficiencies of 89-99% on minus 30 micron dust were calculated. Greater than 95% efficiencies were obtained from the pilot scale tests. The collection efficiency was seen to drop off markedly at face velocities above about 2 ft/sec, and some channeling difficulties were encountered. Because of the operating problems, the full scale installation was shut down after 10 runs. Finally, it should be noted that efficiencies greater than 99% could be obtained in the pilot plant runs by operating with the filter bed temperature at the dew point of the gas being filtered, which introduces a wetted surface effect and further enhanced fines collection.

3. Carnegie-Mellon Crossflow Filter

At present, experimental work on a crossflow filter is also underway at Carnegie-Mellon University. This program, funded by the U. S. Bureau of Mines is under the direction of Professors H. L. Toor and R. L. Zahradnik. Carnegie-Mellon is carrying out the only simultaneous dust-collection-SO_2 sorption study that was discovered in the survey. The filtering medium consists of a slowly moving bed of 1/16 - 1/8" alkalized alumina particles with the solids rate controlled by a vibratory feeder. The gas fed to the test unit, a square shaft one foot on a side with a filter section height also of one foot, consists of heated air with small amounts of SO_2 and flyash added. The gas is loaded with flyash via a small screw feeder. It is recognized that it may be difficult to effect good dispersion of flyash by introducing it in this manner; if the flyash is agglomerated in the gas phase before reaching the filtering section, higher collection efficiencies may be obtained than if the dust particles were completely dispersed.

Two types of experiments have been carried out. The first tests were aimed at determining whether the presence of flyash in the gas or deposited on the sorbent had any effect on the sorption process. With the operating variables as per Table VII, the sorption rate was not affected by the presence of dust in the gas. There was no motion of the solid during these runs nor any attempt to determine dust collection efficiencies.

Data on two further falling solids runs are shown in Table VIII. During these runs, dust collection efficiencies were determined using a final filter designed to trap all particles larger than one micron. No measurable quantity of SO_2 was found in the outlet gas. Note that the 1/16 - 1/8" alkalized alumina particles are considerably smaller than the collection granules used in the Dorfan and Consolidated Coal installations and this might be a significant factor in collection efficiency. These early data indicate that efficient simultaneous removal of both flyash and SO_2 can be achieved in a granular bed device although some question can be raised with regard to possible agglomeration of the flyash. Future effort in this study will be directed toward the establishment of breakthrough conditions for both flyash and SO_2.

B. Intermittent Moving Bed Filters

1. Squires Designs

Dr. A. M. Squires has developed two panel filter concepts in which accumulated flyash is removed periodically through the gas entry louvers. The two designs are called the LS (Loose Surface) and the GSC (Gas-solid Contactor), respectively. Both are essentially fixed bed devices set up in narrow vertical panel arrangements with intermittent movement of solids. The more recent GSC design provides counter-current contacting of the gas and solid (Figure 6). The shaft, 4 inches thick by 12 inches wide, contains four separate types of slatted elements upon which the granular solids are supported (Figure 7). The normal gas flow is from right to left. It is envisioned that relatively coarse granules will be used on the gas exit side to prevent blow-through of the smaller collecting granules. Squires uses the example of 16-30 mesh treating material with 8-12 mesh sealing solid. The internal slats form a number of treating passages within the panel into which the gas is distributed by the right-hand inlet slats. During operation as a sorber, counter-current gas-solid contacting can be achieved by balancing solid and gas flows such that the sorption front always remains within the treating zone. The flyash builds up on the exposed collecting granule surfaces at the gas entry side. When it is desired to remove this buildup or to effect solids movement, the gas flow to the panel is shut off and either

a. A relatively small amount of gas is directed through the panel from the left as a rapid reverse pressure surge for a brief period of time, a technique called puffback, or

b. The right-hand most slats can be struck a sharp upward blow via an automatically controlled hammer system (Figure 8).

It is estimated that the bed cleaning will cause about a four to one removal ratio of collection granules to flyash by weight. Because of the possibility of taking advantage of the counter-current effect, this GSC design is of greater interest as a sorber-collector than the LS configuration. It has never been experimentally tested, however, but it is projected that 99+% collection efficiencies will be achieved at 10 ft/min face velocity, and possibly at 15 ft/min.

The earlier design, the LS contactor (U. S. Patent 3,296,775) (Figure 9) has been the subject of some experimental study as a dust filter along with the puffback bed cleaning technique. This LS configuration is simpler than the GSC in that it contains only one or two types of slatted internal elements (depending on whether screening or a separate support slat is used for solid retention on the clean gas side). Since the panel thickness is only 2-3/8" compared to the 4" for the GSC, lower pressure drops would be expected under the same operating conditions. In the LS design, upslat louvers are used on the dirty gas side as compared to the downflow treating zones in the GSC.

Tests were carried out on the LS design from the standpoints of collection efficiency and of operation of the puffback mode of cleaning. The experimental arrangement is shown in Figure 10. Typical results reported are given in Table IX. Sand was fed to the panel from an overhead bin. Air was pulled through the unit by a blower downstream of the filter. The collection efficiency was estimated from the amount of flyash caught in bag filters located past the blower. Flyash was introduced to the panel by elutriation from a fluidized mixture of sand and flyash. The flyash stripped from the fluid bed is carried by the fluidizing air to a tee where it was combined with the main air supply. The puffback bottle was connected to the clean side of the filter, and the air in the bottle was compressed by a small hand pump. The high collection efficiency results obtained must be considered as preliminary, again because of the possibility of pre-filter flyash agglomeration. In the tests on the puffback cleaning mode, the variables studied included tank pressure and volume, port area of puffback valve, the volume between the puffback valve and filter panel, time of gas surge flow, and slat angles and dimensions. As a result of these studies, a useful range of operating conditions were derived which were applicable to the configuration tested. Tests of this nature would have to be continued if this method of bed cleaning were to be applied to the GSC design.

2. Granger Filter

A concept has been developed by the Granger Filter Company of Wakefield, Massachusetts which can also be described as a type of intermittent fixed bed. The basic configuration is shown in Figure 11. A chamber is divided by a support screen. Granular particles are introduced for a short period of time into the stack gas upstream of the chamber, the size of the granules and the gas velocity being such that the collection granules are transported to the support screen where they form a uniform coating. In this manner the granular layer is set up and flyash collection and sorption proceeds until it is desired to clean the bed and remove the particles. The ducting is arranged so that gas flow can also occur through the filter cell from the opposite direction. The company claims that a reverse flow will very quickly dislodge all solid material from the side on which it was first deposited. This material should thus be quickly removed and the reverse flow of stack gas which transports these solids can be initially diverted to a small secondary collector such as a cyclone or baghouse. Fresh granular material is also introduced at the same time into the reverse gas flow, and a granular filter forms on the other side of the support screen and the cycle is repeated. In an operating unit, a number of these filter chambers would be built into a single device so that only a small fraction of the individual filter segments would require cleaning at one time (the same principle holds for the Squires designs as will be described in Section IVA). The Granger Filter Company markets a liquid filter based on the general concept for use in the purification of industrial water systems. This extension of the idea to gas systems is reasonable but it has not been tested and considerable development would be required.

C. Fixed-Bed Filters

1. Ducon Company Design

The Ducon Company of Mineola, Long Island, New York has had a fixed bed granular filter under proprietary development for several years. In principle, the Ducon device is similar to the Squires concepts. Granular material is confined by means of slatted elements to chambers in the annular region between two metal cylinders. The outer cylinder is made of wire mesh, while the inner cylinder consists of slotted pipe. Thus, an individual module consists of a vertical cylindrical shaft containing a number of separate annular compartments. The dirty gas flows inward and leaves the unit through the inner cylinder. Dust removal from the compartments and the outer screen is effected by a small volume reverse surge of gas from within. The outer screen is sized, however, to prevent any loss of granular material during bed cleaning in contrast to the Squires devices which are designed to remove granules along with the dust. For the Ducon design to be used as a sorber, therefore, regeneration of the sorbent would have to occur in place. The design gas velocity for this device is 50 ft/min. Laboratory data on several materials are shown in Table X. Note that collection efficiencies on relatively coarse dusts such as flyash or fluid cracking catalyst (not shown), although at the 98-99% level, were still lower than those obtained on much finer materials such as the electric furnace iron oxide which was 85% $< 10\mu$ and 60% $< 3\mu$. Evidently the physical characteristics of the dust are another important factor in performance, possibly with regard to agglomeration behavior (cf Section I B). Dimensions of the configuration are regarded as proprietary by the company.

2. Fuller Company Design

The Fuller Company of Catasauqua, Pennsylvania has also had a fixed bed granular filter under proprietary development for several years. Testing has been carried out both in laboratory units and on a pilot scale. The Fuller device is essentially a horizontal fixed bed. Dirty gas is treated in the bed in a downflow manner. Collected dust is removed by means of a periodically applied steady blowback of gas (at again a much lower rate than during the dust collection portion of the cycle) up through the bed which fluidizes and transports the flyash to a secondary collector such as a cyclone or baghouse. With this configuration also, regeneration of a sorbent would have to be carried out in the same equipment as the sorption since the design does not allow for transport of a sorbent material from the bed. The company emphasizes that the blowback technique involves rather vigorous agitation of the collection particles, and it is therefore desirable to employ a granular material having high mechanical strength. The Fuller Company, like Ducon, also does not choose to release any detailed information on the physical and operational characteristics of their filter.

Tests carried out by Fuller on this equipment have given collection efficiencies of over 98%. The performance was found to be strongly dependent on the physical characteristics of the particular granular collecting material used. The company would not reveal, however, what those critical characteristics were.

3. The Lurgi-MB-Gravel Bed Filter

The German company, Lurgi Apparatebau Gesellschaft M.B.H. of Frankfurt, has marketed in Europe in the past a device called the Lurgi-MB-Gravel Bed Filter (Figure 12). As of late 1968, the device was in a period of redesign. In this design, a horizontal bed of granules acts as the filtering medium for an upflowing dirty gas. Periodically, the gas flow is stopped, and the bed, which is mounted on springs, is cleaned by a vibrating mechanism which causes the dust particles to shake loose from the filtering solid and fall through a supporting screen to a collection hopper. For a desired application, it is possible to stack two or more filter beds, one over the other, in a given device, with the bottom-most layer consisting of relatively coarse material to act as a prefilter. Typical operating data are given in Table XI. It should be noted that the intended application for this device was for relatively small gas flows rather than for stack gas service. At last report, the company was considering the redesign of this equipment for larger gas-handling capacities.

III. SORPTION MODELS

The problems involved in prediction of dust collection performance of granular bed devices were discussed in Section 1A. It was also pointed out that no data other than those from Carnegie-Mellon exist for simultaneous SO_2 and flyash removal on a granular device. On the other hand, the problem of modeling the sorption process in a given type of device is far more tractable than that of modeling dust collection provided that reliable rate and capacity data are available for the sorbent considered. In order to determine the probable sorption behavior of the main types of device considered, mathematical models were developed for SO_2 sorption on alkalized alumina for:

a) Steady-state counter-current gas-solid contact (Squires GSC)

b) Steady-state crossflow gas-solid contact (Dorfan type)

c) Unsteady-state fixed bed gas-solid contact (Fuller and Ducon designs)

The form of the solution of problems of this type is highly dependent on the rate expression for the sorption step. In general, models were applied in calculations at face velocities considered reasonable for good dust collection performance. The object was to determine whether there was any problem with regard to 90% SO_2 removal under those conditions.

A. Rate Expression

The rate expression used was an empirical representation of Avco rate data for SO_2 removal on 1/16" alkalized alumina particles at 300°C from a simulated flue gas containing H_2O and NO_x.

The expression is

$$\frac{2.14}{y}\frac{\partial W}{\partial t} = \left(\frac{675}{1 + 6.2S}\right)\left(1 - S^2\right) \qquad (1)$$

The capacity $W_o = 0.32$ $\#SO_2$ / # sorbent where

y = mass fraction SO_2

W = sorbent loading of SO_2, lb SO_2 / lb sorbent

t = time (hrs.)

W_o = sorbent loading at saturation, lb SO_2 / lb sorbent

S = fractional saturation of sorbent ($S = W/W_o$)

This expression gives a faithful representation of the data for values of S between 0.1 and 1.0. It does not show the very high rates observed at low loadings(18), and it is therefore conservative between S = 0 and S = 0.1.

B. Analysis of Squires GSC Contactor

As described in Section II, operation of the GSC device requires adjustment of the solid flow rate so as to confine the sorption front to the counter-current treating zone. Actual operation is of an intermittent nature, and in fact, unsteady-state sorption is the process when the solid is not in motion. It is possible, however, to treat this type of device as a steady-state counter-current moving bed provided that the cycle has a constant period, and that the sorption time in a given cycle is short enough so that the sorption front will not break through the treating zone. It is also assumed that the solid in the treating zone moves as a plug with no backmixing. The analogy is illustrated in Figure 13. Plotted is the relative position of a point originally at the top of the treating zone as a function of time. The sketch suggests that the straight line representing the steady moving bed can satisfactorily approximate the stepwise motion of the intermittent fixed bed. On this basis, a model for a counter-current moving bed sorber was applied to this contractor. The derivation and calculations are presented in Appendix A. The panel dimensions and conditions are given in Figure 6 and Table XII. The panel sizing used was the same as was used in the mechanical layout described in Section IV. In applying the counter-current model to the GSC design, the entire sorbent volume contained between the top of the treating section (Point A in Figure 6) and the right-hand slat (Point B) was taken to form a rectangular plug about 29/64 inches wide by 12 inches deep. This assumption of a constant flow geometry will introduce no problem as long as the sorption path length required for a given extent of SO_2 removal is less than the length of the sorption section. If this condition does not hold, then the simulation of the intermittent fixed bed by the steady counter-current model is not valid either.

Calculation using the condition in Table XII gives a required sorption path length of 0.027 feet, well under the value of 0.341 feet available. Operation of the GSC device under these conditions is therefore not critical with regard to sorption. Indeed, the counter-current contacting concept is not efficiently utilized at this relatively low gas velocity specified for dust collection. The pressure drop through the bed was estimated at about 3 inches w.g., not including any ΔP contribution from flyash buildup.

Since the counter-current contacting potential of the GSC device is not effectively used at the gas and solid-flow rates chosen for the above example, a further calculation was undertaken in order to determine a set of conditions when the GSC is actually operating as a counter-current sorber. Note that this will require a higher gas velocity to the unit, and it is questionable as to whether good dust collection performance would be achieved. For this calculation, it was assumed that S, the outlet fractional sorbent saturation was 50%, and that sorption did occur throughout the 0.341 foot path length instead of the 0.027 foot length obtained in the previous calculation. The results led to a value of G of 966 #/hr ft^2 corresponding to a face velocity of 130 ft/min and an estimated pressure drop of 39.6 in w.g.

C. Crossflow Contactors

A model was developed for a falling solid-crossflow gas contactor in order to check the effects of operating variables on SO_2 sorption in such a configuration. Calculations using this model are applicable to the Dorfan and Carnegie-Mellon contactors. The bases of the derivation are shown in Figure 14. Regenerated sorbent enters the apparatus at the top flowing at a rate of L lbs/hr-ft^2 into area A_S. Gas at rate G lbs/hr-ft^2 flows from left to right into area A_G. In the contactor, concentration gradients are set up in both the gas and solid phases and both will exhibit non-uniform exit profiles. The governing equations are given in Appendix B. The conditions used for application of the model are given in Table XIII. These calculations indicated essentially complete removal of SO_2 and an average outlet solids loading of 0.137 lb SO_2/lb sorbent. The SO_2 profile in the bed indicated that very little sulfur penetrated beyond 60% of the thickness. The estimated pressure drop was 7 inches of water. Thus, there is no criticality with respect to sorption in the crossflow contactor operated at dust collection conditions.

D. Fixed Bed Contactors

The analysis of the steady-state crossflow system led to a model having two independent distance variables. The unsteady-state fixed bed also has two independent variables, one distance and one time, and the solution is worked out in a similar manner to the crossflow case (see Appendix C). The conditions employed for the fixed bed calculation are given in Table XIV. The results show that after the bed has been operated for 5.5 hours under the above conditions, the sorbent at the top of the bed is loaded to less than 0.2% of saturation and, likewise, only about 0.2% of the inlet SO_2 is leaving the sorber in the outlet gas, i.e., no appreciable (defined as 10% of inlet SO_2) breakthrough has occurred after 5.5 hours of operation. The calculation indicates that a fixed bed sorber can be operated in place for a number of hours before regeneration is required. It is likely that a number of flyash removal cycles will be necessary before the sorbent needs regeneration. Thus, sorption does not appear to be a critical problem in a fixed bed contactor either, similar to the results for the counter-flow and crossflow beds.

E. Summary

For all three designs for which calculations were carried out, the results showed that SO_2 sorption on alkalized alumina does not limit the capacity of the given device at gas velocities stipulated for dust collection.

In order to more fully take advantage of the counter-current contacting capability of the GSC design, the extimates indicated that the inlet gas rate could be increased by a factor of nine. Calculations for the crossflow device indicated that all sulfur was adsorbed in about only 60% of the bed thickness. The fixed bed calculations indicated SO_2 breakthrough would not take place for over 5 hours.

Although these devices could be used more efficiently for sorption at higher velocities (but at increased pressure drops), it is uncertain how seriously the filtration behavior would be affected.

IV. MECHANICAL LAYOUTS

The preceding sections contain descriptions of various granular bed devices, and of experiments and calculations carried out with regard to them. Owing to the scarcity of detailed experimental data, however, it is not possible to arrive at unequivocal statements as to their respective technical merit. Enough information does exist to permit some preliminary cost estimates to be obtained to serve as the basis of assessment of the economic potential of these designs. The estimates must be based, however, on projected arrangements of this equipment and auxiliaries for a power station application.

Of all the devices considered, the two major concepts not presently being commercially pursued were the Squires panels and the crossflow panels. Preliminary mechanical layouts were worked up for these designs to serve as bases for the cost estimates. Not enough information was made available by the Ducon and Fuller companies as to the exact dimensions of their respective filter modules to permit layouts to be developed for these designs. In those two cases, it was necessary to rely solely on information received from the two companies.

The layouts were based on treating stack gas from an 800 mw power station at a sorption temperature of 300°C. The actual gas volume projected was about 3×10^6 cfm.

A. Plant Layout for GSC Design

In order to effectively utilize a concept such as the GSC contactor for an industrial application, it is necessary to arrange the individual panels in a way that is compact with regard to space. The layout presented below is largely based on material presented by Prof. Squires in an unpublished memorandum on the design of LS panels. Essentially the same configurations have been used for the GSC contactor with appropriate changes in dimensions. A row of panels, each 20' x 1' x 4", consists of 32 hung end to end with gas passages, denoted clean and dirty gas aisles, as shown in Figure 15. In one box, 12 such rows are suspended, making the box dimensions 16' x 12'. The face velocity used to the panels was 15 ft/min. One dirty gas aisle feeds two panels or a total area of 40 ft^2; this area will thus handle 600 ft^3/min. There are 16 of these double panel arrangements in a row and 12 rows in a box giving an overall capacity of 115,200 cfm/box. The gas volume estimated from the power station can be treated in 25 such boxes.

In order to clean flyash from the panels a puffback system, external to the contacting panels, is built into each box so that the required reverse pressure surge of gas through the panels can be effected. A diagram of a single box is shown in Figure 16.

The flyash-granule mixture dropping from the panels falls into a lower bin from which it is moved by a screw conveyor (or other means) outside of the box to a pneumatic line. The solid is then conveyed pneumatically by air at about 100 cfm in a 3 inch line to the top of the box to a granule-flyash separator which simply consists of a conveying line opening up into a 12 inch pipe. The gas at reduced velocity in the larger pipe will not be able to carry upward the larger collection particles but will still be above the terminal velocity of the flyash and attrited granular material. The granular material separated out will then be conveyed over the tops of the boxes to a regenerator if a sorbing granule is used, or returned to the overhead feed bin directly if the device is used solely as a dust collector. The conveying air carrying the flyash and attrites is directed to a flyash collection and removal system comprising one or several small cyclones and a central collection tank from which it is removed from the plant. The exhaust gas from the flyash collection cyclones will be returned to the dirty gas side of one of the boxes.

Sorbent will be returned to each bin from the regenerator via horizontal lines again located at the tops of the boxes. The mode of conveying the sorbent to the regenerator and back can be a pneumatic high ratio solids-gas flow technique. A small cyclone located at the top of the storage bin should serve to collect all the sorbent conveyed in this manner and can be vented to the atmosphere. It should be noted that if sorbent will be sent for regeneration it is not necessary to carry out the granule-flyash separation at the top of the box, but if this were more convenient this can be done at the side of the box closer to the ground level. Since the regenerator itself will most likely be a moving bed device with the solid continuously flowing downward, however, a means of raising the sorbent to the top of the regenerating tower would be required in any event. For this reason, the arrange shown in Figure 16 is considered as reasonable. A rotary valve at the bottom of the sorbent collection cyclone controls the return rates of granules to the overhead bin.

The diagram depicts a number of distribution lines for spreading the returned solid over the entire area of the overhead hopper.

The dirty gas enters at the lower portion of the box just above the lower bin, and the clean gas exits above the panels at the top. Regenerated granules travel by gravity to "clean sand" compartments located just above the panel openings by means of one-inch return pipes from the overhead sand bin. These pipes pass through the puffback ductwork. Clean gas flows into headers located above the clean gas aisles. Normally open dampers at the tops of these headers permit free passage of this gas to the exit ductwork. When a particular row of panels is to be puffed back, however, the dampers on the clean gas headers in that row are closed and puffback gas enters each clean gas aisle by means of four inch pipes dumping into the header (except for the first and last headers in a row which are only two inches wide). When a row is shut off for puffback, the gas flow distributes among the remaining eleven rows in the box. It is estimated that the overall puffback time requirement is about 30 seconds.

The puffback system as shown in Figure 16 consists, to start with, of two 18" diameter by 7 foot high puffback tanks per box, each serving one-half of a given row. Situated at the top of the puffback tank is the main puffback valve. According to Squires this valve must be capable of opening in less than 0.1 second and must have a port opening of 0.125 in^2/ft^2 panel or 40 in^2 for half a row of panels (20 ft^2/panel x 16 panels). No valve meeting these requirements is commercially available at the present time. To accomplish the rapid opening, one Squires proposal is a piston valve normally seated in the top of the puffback tank. The valve, shown in Figure 16 extends about 7 inches below the top of the tank and is held in place by a mechanical stop. When puffback is to occur the tank is pressurized by an external air compressor to 60 psig. The stop is then pulled back and the tank pressure forces the piston upward. By the time the piston completely clears the 18" segment and permits gas to flow from the tank, it should be moving at a very high velocity and give a proper puffback reaction. An upper stop catches the piston and it drops back into the tank for another cycle. The gas passing through the puffback valves dumps into a feeder duct 10" x 10", located in the middle of the 12 rows and running for half of the 16 foot length of the box. Six 10" x 10" ducts run to the individual half rows on either side of the feeder duct; these lines feed into a 10" x 12" duct running parallel to each row and directly above it. It is from these 10" x 12" ducts that the 4" pipes lead to each individual clean gas header in the half row. It is also through these same 10" x 12" ducts that the 1" clean sand return pipes pass. The puffback gas is diverted to the particular row of interest by means of damper valves placed in each of the 12 lines running from the central feeder duct. These externally activated dampers need not be fast acting since they can be set before the puffback valve opens.

Some preliminary calculations were made with regard to the pressure drop relationships occurring during puffback. These were mainly directed at determining whether any relatively large pressure surges would be experienced back at the boiler as a result of rows being valved in and out of the system for puffback. Frequency of puffbacks required with dust buildup were very crudely based on Squires estimates for the LS design of a maximum allowable dust buildup of 2 oz/ft^2 panel area. A higher bed dust loading would lead to poor flow distribution. Another figure used was 0.57 in w.g./oz dust/ft^2 panel added pressure drop caused by dust buildup over the ΔP caused by the clean gas flow alone. Note that these are probably not good values for use with the GSC configuration, but are only used in this case for a very rough estimate. For a situation where the flyash concentration in the hot gas is 1.5 gr/ft^3, any given row, on this basis, should be puffedback about every 40 minutes or the puffback valve dumped every 3.3 minutes. This leads to an estimate of 3.3 in w.g. through all panels before puffback, 3.6 inches during puffback and 3.2 inches just after puffback. Just before a given row is puffedback, the gas flow rate through it is about 30% less for the same ΔP than through the row just previously cleaned. As a result of these calculations it would appear that no mainline pressure difficulties are expected with regard to the puffback switching.

The sorbent circulation rate from a single box was estimated to be 5940 lbs/hr. The overhead and lower granule bins were sized to allow at least a 15 minute holdup.

The boxes for this arrangement were assembled in two rows of 13 with the ductwork running through two boxes in separate rows (Figure 17). Three foot spacing between boxes was used. Running the gas for two boxes into the first box at about 30 ft/sec requires an opening of 11 feet above the collection bin and below the panels. The aim is to achieve an even distribution of gas by providing a $5\frac{1}{2}$ foot high opening at the back side of the box for dirty gas flow to the next row, so that just one-half of the entering dirty gas will be treated in the first box. One problem that arises, however, is that the dirty gas that passes through will probably entrain some of the solid material that falls down during puffback. A method of counteracting the entrainment is indicated in Figure 16, by simply showing a screen extending in front of the dirty gas exit from the first box, and an inclined duct leading to the $5\frac{1}{2}$ foot high opening at the front of the second box where the purpose of the latter is to prevent accumulation of solids in the ductwork between boxes (Figure 17). Finally, the clean gas exit duct from the second row box is eleven feet high.

B. Plant Layout for Crossflow Contactor

The shaft modules for this layout were arranged in the same manner as was used in the Consolidation Coal Filter (Figure 4) to gain a compact form. Six separate filter shafts in parallel were enclosed in a single box. The individual shafts were taken as nine feet wide, twenty feet high, and six inches thick. Although the largest single Dorfan shaft arrangement as mentioned previously was 9' x 14' x 1", the assumption was made for this layout the height could be increased to twenty feet while still maintaining good gas distribution. The thickness was reduced from 1 foot to 1/2 foot, and the design velocity used was 1-1/2 ft/sec. The six shafts were arranged in two rows of 3 on either side of a central dirty gas duct. Clean gas ducts run on the opposite sides of each of the two rows (Figure 18). Each shaft presents a filtering area of 180 ft^2 and, at a face velocity of 1.5 ft/sec, can handle 16,200 cfm. Each box then will take 6 x 16,200 = 97,200 cfm giving a requirement of 30 boxes for the service. If the inlet dirty gas duct velocity is 30 ft/sec, an inlet duct width of about 3 feet will be sufficient. Likewise, the two outlet ducts from a single box will each be 1-1/2 feet wide to accommodate the clean gas at the same velocity, making the width of the box ktself 7 feet. From a practical standpoint, the interior duct work in an actual installation would have to be most carefully designed to insure even gas distribution to all six shafts. For example, it is likely that the inlet gas duct will be tapered inward as it proceeds through the box and the outlet ducts tapered outward so as to allow for the different total gas flows which will be present in the respective inlet and collection ducts, at different positions within the box.

The above discussion and Figure 22 are based on a given box only handling gas for itself alone. If the boxes are arranged, as would be preferable, in two rows of 15 similar to the GSC layout, gas flow for two boxes would be again, run into the first. This would require enlarging the ducting dimensions of a single enclosure to handle the increased gas flow. Effectively, the overall box width would be increased from 7 to 10 feet as shown in Figure 19.

It was further assumed that the residence time of solids in the bed was 200 minutes which gives a solids flow rate of 7,290 lb/hr for a single box. The overhead and lower granule bins were again sized on the basis of at least a 15 minute sorbent holdup.

The granule flow rate is controlled by four rotary solid control valves, one located at the bottom of each shaft, on each side. The valves run for a length of 13-1/2 feet. From the rotary valves the flyash-granule mixture falls into the lower bin from which it is conveyed to a pneumatic line and then to a granule-flyash separator. The solids treatment is exactly the same as in the GSC layout.

V. PRELIMINARY COST ESTIMATES

The cost estimates obtained were based on the layouts presented in the preceding section. The procedure used was to compile a list of major equipment items and then to obtain estimates for the individual entries. The total was then multiplied by a factor of 3.1 in order to approximate items such as installation, construction, contractor's fee, contingency, and working capital costs, thus giving a value for the total capital investment. This is the same procedure as was followed in Reference 18. Since the most important items in the cost estimates, viz., the panels in which the active granular material resides during contacting, are non-standard fabricated units, assistance in estimating this cost was sought from several metal fabricating companies. Several of the other items were also costed based on vendor information and a complete list of the companies who assisted in the analysis appears at the end of this section. The general approach was to attempt to obtain realistic costs which reflected economies resulting from the scale of fabrication required for a facility serving an 800 mw power station. Costs for the necessary auxiliary equipment for an alkalized alumina sorption process, i.e., regeneration equipment, gas producers, and Claus units were taken from Reference 18. The costs reported are thus from the standpoint of a power plant installing this equipment rather than from a company selling the devices. Operating costs were divided into four categories: a) labor, maintenance, and overhead, b) raw materials, 4) capital charges, and d) utilities. The data for the first category was also largely based on the work described in Reference 18. Finally, the cost estimates are presented on the basis of a combination dust-collector - alkalized alumina SO_2 sorber installation in both new and existing power plants, and as a dust collector alone for a new installation. In the latter case, the granular material is taken to be graded sand instead of the more expensive sorbent.

A. Squires Contactors - Dust Collector-sorbers and Dust Collectors

By far the most important equipment cost item for these devices is the panels for gas-solid contacting. The position was taken that these panels would be produced in quantity by a fabricator and then assembled by the contractor on-site during construction, i.e., they would be regarded as delivered equipment to be installed. For the GSC contactor each panel is made up of four separate types of elements. For the 26 enclosures required for an 800 mw facility, about 5 to 6 million separate pieces would be required, indicating the desirability of a mass production operation. This is based on panel dimensions of 20 feet by 12 inches by 4 inches. Each panel requires two side plates over the 20 foot by 4 inch dimension. In order to reduce assembly costs, it was decided to have these side plates slotted to receive and support the individual elements, and the internal vertical bearing bars shown in Figure 7 were eliminated. The elements were taken to be fabricated out of 20 gauge cold rolled stock and the side plates out of 16 gauge material. This leads to an individual panel weight of 309 pounds and an overall estimated cost of $182.35 or $0.59 per pound, a composite figure arrived at by combining separate fabrication and assembly costs. No single company was interested in estimating the costs of complete panels; several cited inability to handle either the large number of pieces or the large height of panel assembly. About 10,000 panels are required for a 26 enclosure installation and this leads to an equipment cost of $1,823,500.

Another large single cost item is the ducting required between enclosures. Normally, this item could be considered under the category of piping and thus be included in the factor multiplying the delivered equipment cost. In the cases under consideration, however, since this is such a large item and relatively insensitive to the design concept, it was decided to include it as a separate entry, and make an appropriate reduction in the multiplicative factor employed. On this basis, the factor was reduced from 3.5 as was used in Reference 18 to 3.1 for the present study. The ducting estimate for the Squires designs was \$300,000. Also required to be incorporated into the ductwork is a damper system which will permit the isolation and bypassing of any individual enclosure or row of two. It will be thus possible to isolate either the entire complex during startup or shutdown periods or any individual row for maintenance. The cost of this damper system was estimated at \$270,000 for the GSC design.

With this background on the main items, capital cost estimates are presented in Table XV for the Squires contactors both as sorber-dust collectors and dust collectors alone under the conditions indicated. The other auxiliary items required are each relatively small and their estimation is straightforward. Some comment should be made, however, regarding the overall power plant blower. It was arbitrarily assumed that the plant pressure drop exclusive of any sorption- dust collection equipment was 8 inches w.g. For the Squires GSC device, provision should be made for an additional drop of about 4 inches. Assuming further that the plant uses i.d. blowers rated at 350,000 cfm and handling gas at about 300°F, 6 such units would be required. If a GSC unit were to be installed in an existing power station, it was assumed that the original blowers could be modified to allow for a higher drop by changing the size of the fan wheel at an estimated cost of \$2,000 per unit. In a new installation, the blower cost was estimated as the difference in investment required with and without GSC equipment. The values for an LS installation were estimated in part from figures developed by A. M. Squires. As a dust collector alone, it is obviously preferable to the more complex and expensive GSC arrangement.

In all the Squires cases, the costs included estimates for a puffback cleaning system for the bed. Since it is recognized that this mode of cleaning involves some fairly complicated ducting and also relies on the development of a very fast-acting valve, it may be desirable to consider a mechanical cleaning system to take the place of the puffback equipment. Squires has set forth the concept of an automatically controlled hammer system which will act to periodically strike a sharp blow to the slotted elements on which the dust actually collects. If one such hammer system can serve two panels, as was indicated in Figure 8, then 192 x 26 = 4992 such devices would be required. If the mechanisms and actuators sold for about \$25 each, this would require an investment of about \$125,000 delivered. This is in contrast to an estimated reduction in investment of about \$267,000 delivered if the puffback system costs are not included. Some reduction in capital investment might thus be expected if a mechanical cleaning system were to replace the puffback piping, especially if the cost of the mechanical components could be kept very low. This is no comment on the relative technical merits of the two modes of cleaning.

With regard to the dust collector estimates, it should be noted that although these units were costed for the same gas volume as would be found in the sorber cases, they could be operated at a lower temperature and thus handle less gas. The most important figure for comparison of dust collection equipment is thus the cost per cfm rather than per kw.

Operating costs for the Squires devices are shown in Table XVI. A 14% yearly charge was taken on the capital investment to include depreciation, taxes, insurance, and return on investment. Labor, maintenance, and overhead costs were estimated from information presented in Reference 18. Raw materials costs estimates were based on a filter medium attrition rate of 0.1%. Each 0.1% increase in attrition rate corresponds to $0.10/ton additional operating cost.

Utility costs were estimated from power requirements for the individual auxiliary compressors found in the system and from a prorating of that part of the overall power requirement needed to move flue gas through the equipment. The operating costs shown do not take a credit for the sulfur recovered from the Claus unit.

B. Crossflow Contactors

Capital and operating cost estimates for crossflow gas-falling solid shaft contactors are presented in Tables XVII and XVIII.

The major differences that the crossflow design estimates show as compared to the capital costs for the Squires devices reside in the panel costs and the absence of a separate bed cleaning system.

Since there are many fewer panel sections required in the crossflow contactors owing to the higher filter face velocities used, the cost estimate for this item is far lower than in the Squires designs. The panels in this case consist of louvered sheets, 20 feet by 9 feet, and there are 12 such sheets required in a single enclosure. The louver openings are about 1/2 inch, and the sheet is made of 16 gauge material. The unit price of $0.59/pound as was used for the Squires panels was applied to these sections as well.

The pressure drop through the beds was estimated here to be about 7 in w.g., a figure considered too high to permit modification of blowers in an existing power station. This would require complete replacement of the equipment in an old installation making this entry higher than the corresponding blower cost item for an all new installation. In the latter case, the blower cost attributable to the granular bed equipment was obtained from the difference between equipment required to overcome a 16 inch ΔP and an 8 inch ΔP.

C. Squires Contactors - Sorbers

It should be recognized that the Squires and crossflow estimates are not based on the same superficial velocity but on values a factor of six apart. This discrepancy was based on information supplied by the developers of these concepts, and although no results exist from which an experimental estimate of maximum face velocity can be obtained, it is of interest to speculate as to the effect on the costs of permitting the face velocity in the GSC or LS devices to increase to, say, 1.0 ft/sec. In order to illustrate this effect an estimate was made for a GSC system operating at 1.0 ft/sec. Capital costs were obtained of $7.09 per/kw and operating costs of $1.14 per ton coal, values substantially below the previous GSC estimates.

In addition, a cost estimate was made for the maximum sorption velocity case for countercurrent contacting, discussed in Section IV. Only three boxes would be needed to handle the total gas flow under these conditions. This led to estimates of total capital investment of $7.19 per kw and an operating cost of $1.35 per ton of coal. The latter value for a 3 enclosure system is not very different from those for the 26 box installation owing to the high power requirements.

A comparison of these two estimates indicates that a velocity of 1.0 ft/sec is nearer optimum from a cost standpoint than the "maximum" sorption velocity.

D. Ducon and Fuller Designs

Estimates were received from the Ducon Company for prices of systems using their design. In order to facilitate comparison with the Squires and crossflow systems, the following case was chosen:

Thirty-five Expandable Bed Filter Units, 12 feet diameter by 35 feet high with both vessels and internal elements fabricated from carbon steel. Each unit contains 85 filter elements 1' - 0" O.D. x 20 feet long. Price FOB shop -- approximately $2,500,000 O.D.

Instrumentation and controls including backwash air piping - $100,000.

Estimated maximum pressure drop - 8 inches w.g.

Note that the above estimate does not include the cost of intervessel ductwork. Although the Ducon design is presently oriented toward dust collection alone, an estimate was also made for a sorption system using this design consisting of two complete sets of units. One set would be regenerated while the other was used in sorption-dust collection service.

Since the Ducon figures are for equipment costs, the total capital investments were estimated by again applying the factor of 3.1.

The Fuller Company declined to submit a cost estimate for their design. They have, however, submitted a plan view of a proposed layout for the required service (Figure 20). A collection efficiency of over 99% was anticipated with an overall pressure drop of less than 10 inches w.g. Based on the dimensions given in the sketch, the volume of gas handled per unit of plot area (not including the stacks) is about 230 cfm/ft^2 plot area.

E. Comparative Cost Estimates

Values of capital cost per kw and per cfm of gas handled and operating costs per ton of coal are presented for all the major processes in Table XIX and indicated graphically in Figure 21.

Published figures for other SO_2 removal processes are given in Table 20. The estimated costs for the granular devices are not out of line with estimates for other processes.

Also included in Table XIX were estimates for electrostatic precipitation equipment obtained from manufacturers. Typical estimates ranged from \$0.60 to \$0.70 per cfm for a 99% efficient unit in the required size range. For dust collection equipment, then, the lowest estimate obtained was for the crossflow design, followed by electrostatic equipment, the LS, Ducon, and GSC designs.

Calculations indicate that the overall gas treating capacity of the Squires units are about 200 cfm/ft^2 plot area, and for the crossflow design, about 100 cfm/ft^2. These figures fall into the range of plot area requirements for electrostatic precipitation equipment. There does not, therefore, appear to be any great space saving advantages that would accrue by using granular bed equipment. The Squires contactors appear to offer more efficient space utilization than the crossflow, but this will be established when the maximum allowable velocity for a given type of service, either sorption or dust collection, is determined experimentally.

The cost estimates obtained do not predict a very wide cost spread between the types of equipment analyzed. To be sure, the differences in capital investment between the Squires and crossflow designs, and the even more pronounced case of the Ducon fixed bed in the case of sorber-collectors, and between the crossflow and GSC designs as dust collectors are regarded as real, but as was shown, these costs strongly reflect the operating conditions chosen for the economic evaluation. The particular conditions used may not be near optimum, and indeed, any of these devices may be proven technically unsound in practical operation. On the other hand, the economic analysis brings out that:

a) Cost estimates for sorption-collection equipment compare well with the estimates in Table 20 for SO_2 removal techniques actively being considered at present and offer the added advantage of flyash collection at the same time.

b) Cost estimates for equipment for dust collection alone compare favorably with electrostatic precipitator costs.

One may further consider the case where an "optimum" GSC sorption system, i.e., at a face velocity of 1.0 ft/sec, is employed in conjunction with an electrostatic precipitator. This assumes that there will be significant flyash carry-over from the GSC when operated at the higher face velocity. Data from Table XIX give combined values of \$12.62 per kw for the capital cost, assuming an electrostatic precipitator at \$0.70 /cfm operating on gas at 300°F, i.e., the total gas volume handled by the dust collector is 2×10^6 cfm rather than 3×10^6 cfm. The combined operating cost estimate for a 90% load factor is \$1.65 per ton. These figures are also comparable with the values given in Table XX and the GSC sorber-electrostatic precipitator combination appears to be another reasonable approach to the SO_2 flyash removal problem. It will require a lower capital investment than a GSC sorber-collector system but will have about 10% higher operating costs.

The conclusion drawn is that although the granular bed concept has not been the subject of extensive development effort, it has the potential to make a significant contribution to air pollution control technology. Only if devices of this type are adequately tested can the measure of that potential be determined.

potential be determined.

F. Acknowledgement

The following companies graciously contributed information on equipment and values of equipment and fabrication costs which were useful in compiling the above estimates. Avco is grateful for their assistance to the program. The companies are:

Arrow Louver and Damper Corp., Maspeth, New York

Carter-Day Company, Boston,Massachusetts office

Eastern Tool and Stamping Company, Inc., Saugus, Massachusetts

Johnson Service Company, Boston, Massachusetts office

Link-Belt Company, Boston, Massachusetts office

New England Engineering Co., Medford, Massachusetts

Peter Forg Mfg. Company, Somerville, Massachusetts

Sturtevant Division, Westinghouse Electric Company,
Boston, Massachusetts office

Uno Louver Company, Malden, Massachusetts

In addition to the above, Mr. George Saunders, former chief estimator (retired) for Stone and Webster Company, Boston, Massachusetts, contributed two useful suggestions.

Special thanks are owed to Prof. A. M. Squires for use of his complete file on the LS and GSC devices and also to Messrs. M. I. Dorfan and J. Apt with regard to fabrication drawings supplied and discussion of the Impingo Filter.

VI. DISCUSSION OF FURTHER EXPERIMENTS

A. Ducon and Fuller Designs

Relatively little information is available as to the details of these configurations, and very little information is available concerning experimental tests on these proprietary devices. Although they both may well have great promise, since details of neither the designs nor the previous experiments are known, there is no basis for specifying continued experimental programs. No recommendations can be made for further study as long as these designs remain under complete proprietary control.

B. Crossflow and Squires Designs

Not enough reliable data exist with regard to either the crossflow or Squires contactors that would enable a realistic technical evaluation of these devices to be carried out. From the results of the preliminary economic analysis, as well, there is relatively little basis to choose between them. Using the design velocities for dust collection, the crossflow unit gives lower costs, but if the gas rate in the Squires device could be increased while retaining good dust collection performance, the GSC contactor would appear more economical. The determination as to whether any one concept is actually more attractive will ultimately depend on the demonstration of an efficient and operable device. Only when the particular mechanical problems associated with a given design are able to be solved will a configuration be available that can be costed with assurance. In other words, although a design may appear economically competitive, its true economic value can be determined only when the two questions of efficiency and operability have been experimentally evaluated. A device that does not bear the promise of long term operation after being experimentally tested should not be included in a comparative cost estimate. On this basis, it is recommended that tests be carried out on both the Squires and crossflow configurations. The ultimate objective of such a study would be the demonstration of long term operability of not only the particular main filtering system, but also the required auxiliaries such as the flyash-granule separator. The system design would have to ensure even gas distribution to all the filtering panels and relatively uniform pressure drop over the height of any one panel section.

For the crossflow system, an experimental program may be carried out as follows:

The first phase of the study would have as its object the unequivocal determination of the filtering efficiency behavior of the concept. A single shaft unit would be built with a filtering area at least 10 feet in height and 5 feet in width to test gas distribution and pressure drop effects. Graded sand would be the preferable collecting medium initially and various size ranges can be tested. The unit would be equipped with an appropriate solids control system to ensure steady feeding. The flyash carried into the device should be in dispersed form. A good technique for achieving this would have to be established as part of the study; another possibility is setting up of the unit at a power station and bleeding actual stack gas through the filter for some of the runs. Solid rates, gas rates, flyash loading, dust particle size, granule size, and gas temperature would be among the variables studied. Measurements of the outlet particle size distribution as well as overall loading would be valuable. The louver angles and spacing should be the subject of some preliminary tests to determine a design that will prevent the granular material from being blown from the bed by the gas passing through, and at the same time, give steady downward granule motion without hangup problems. Local velocity traverses should be made across the exit bed surface to determine flow distribution over the height and width as a function of operating conditions, and these tests may lead to several different inlet ducting schemes being tested as well. It may further be found of value to include bed thickness as a variable or to test the efficiency of a unit consisting of two shafts in series as per the direct Dorfan concept.

This study would establish the technical merit and performance characteristics of this type of device. If the results were encouraging, the next phase of the work should involve:

1. Detailed mechanical designs and layouts for assembly of individual shafts in an enclosure and arrangements of enclosures and auxiliary equipment.

2. Economic estimates for the revised layouts as they would apply to actual installations.

3. Construction of an enclosure and testing of its operation.

The undertaking of these latter three aspects would be contingent on satisfactory results being obtained in the first phase of this study.

For tests on the Squires system, both dust collection-sorption characteristics and bed cleaning behavior must be demonstrated, initially on a small scale. Development of an effective and relatively compact bed cleaning system will probably involve the study of several configurations with regard to considering both puffback and mechanical means of cleaning. Test sections of about one foot in height can be used to establish correlations for an idealized performance level which can be a basis of comparison for more complex units. Data should be taken over a range of variables such as were mentioned for the crossflow design.

When the performance and bed cleaning experiments have yielded satisfactory results on short individual test sections, it would be advantageous to proceed to a fairly large (about 10 feet high) dual panel enclosure to emphasize gas distribution, solids feeding, and other operational effects. This enclosure would be analogous to the 10 feet high crossflow section recommended above, since in the Squires case a means of bed cleaning must be provided external to the panel, and the configuration of these means must realistically be bound up with the arrangement of panels in an enclosure. Note that the foregoing experiments are still prior to a detailed mechanical design for a multipanel module.

If the results were encouraging at this point, then the next phase of the effort would involve detailed designs, cost estimates, and large scale experiments, but as in the crossflow case, these aspects of the study would be contingent on satisfactory results in the early tests.

Finally, although the above recommendations have proceeded from the assumption that the granular device will be used in dust collection service and possibly also in sorption service, it is to be recalled that calculations for the GSC panels indicate that significantly lower costs could result from the higher velocities which sorption operation only may permit (Section V-E), and this may be attractive for an installation with an existing electrostatic precipitation unit. Development for this type of application, however, would not change the basic types of experiment to be carried out. A well-functioning solids transport system for the panels would still be essential. Good gas distribution and absence of channeling are still important over a range of flow rates. The specific experiments that could be avoided would be those that pertained to flyash collection efficiency only, admittedly a major task, but nontheless the equipment must be shown capable of good mechanical operation even on a non-particulate - containing gas.

VII. SUMMARY OF CONCLUSIONS AND RECOMMENDATIONS

A. The limited experimental work on granular bed filters that has been carried out is insufficient for a definite technical evaluation of this type of device as a flyash collector, although high efficiencies were obtained under certain test conditions. In addition, realistic estimates of filtration behavior cannot be obtained from theoretical considerations at present.

B. Calculations for the sorption behavior of SO_2 on alkalized alumina in counterflow and crossflow gas moving beds and in fixed beds show that sorption will not be the limiting factor in operation. The maximum gas velocity to a given filter section will be governed by either pressure drop requirements or filtration performance.

C. The development of preliminary mechanical designs indicates that the gas handling capacities per unit of plot area of the granular bed devices considered are of the same order as the capacity of electrostatic precipitation units.

D. Cost estimates for equipment for simultaneous SO_2 sorption-flyash collection compare well with estimates for other SO removal techniques presently being considered.

E. Cost estimates for flyash collection equipment compare well with electrostatic precipitator costs.

F. It is recommended that experimental work be undertaken on individual panels of the GSC and crossflow contactor designs with the aims of:

1. Determining filtration and sorption performance
2. Evaluating the problems of solid motion and gas distribution to the equipment. In carrying out filtration experiments, great care must be taken with regard to the effects of dust particle size distribution and agglomeration.

G. If these experiments give encouraging results, detailed mechanical designs should be drawn up for an enclosure consisting of several panels, and experiments on this "filter box" should be carried out.

VIII APPENDIX

A. Filtration Mechanisms and Correlations

Several mechanisms may be simultaneously operative for particle collection in granular beds. In general, for collection of a particle of a given size, the relative importance of a number of these mechanisms will vary with the gas velocity to the filter.* Diffusional collection, for example, results from random Brownian motion of a particle in the vicinity of a collecting body, and is most efficient at low velocities. In direct interception, on the other hand, the particle follows the gas streamline faithfully and is only collected when it comes within a distance of its own radius to a collection surface (Figure 22). Inertial impingement, controlling at higher velocities, results when the momentum of the particle is great enough to cause it to break away from the streamline as the gas flow diverges around the collector. The particle thus travels closer to the collector enhancing its chances for interception (Figure 23).

A qualitative illustration of how these three mechanisms are displayed is shown in Figure 24, a plot of efficiency against gas velocity. At low velocities, diffusion will predominate, but as the velocity increases, the residence time of a particle in the vicinity of the collector decreases, and with it the likelihood of contact by random motion, and so the efficiency drops. At intermediate velocities the curve flattens as direct interception becomes more important, and with further increase in the gas rate, inertial impingement begins to control causing the efficiency to rise.

Investigators in the field have developed mathematical models for the individual mechanisms, and solutions can be obtained numerically.(9) The models apply to the case of collection on a single body such as a sphere or a cylinder. When more than one collection mechanism is significant, the theoretical problem becomes greatly complicated, and solutions have been obtained only when one mechanism is predominant. In addition, rigorous solutions are not available in a situation where a realistic array of collectors, such as in a packed bed or fiber mat, is considered where the flow pattern around any one collector is influenced by the presence of the others.(10) Furthermore, the theoretical studies have not taken into account other potentially important factors such as turbulence in the gas stream, particle interactions in the gas, loading of the collector with particulate matter previously deposited, and re-entrainment of collected particles.

For these reasons, the greatest practical use for purposes of this study that can be made of existing theoretical or modeling equations for aerosol filtration relates to the specification of dimensionless groups which give a rational basis for correlation of experimental data. This approach has been followed with regard to fiber mat filtration studies (11) (15) carried out on dilute gas streams of essentially monodisperse aerosols with a size range of 0.3 to 1.5 μ in the different experiments. The collection fibers in these studies had individual diameters ranging from 1 to 50 microns in size.

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* For theoretical purposes, collection is defined as the contacting of a gas-borne particle with a collecting surface. It is assumed that this contact will produce an irreversible adhesion, and the particle will not be re-entrained.

For a given filter bed, aerosol, and gas, a dimensionless correlation based on diffusion, flow line interception, and inertial impingement mechanisms and neglecting any effects of bed loading or dust particle interaction can typically take the form

$$\eta = \varphi\left[N_{Si},\ N_{Sf} \right] \qquad (1)$$

where η is collection efficiency on weight fraction of dust removed;

$\dfrac{k_m \rho_p D_p^2 U_o}{18 D_B \mu} = N_{Si}$ is an inertial impingement parameter, and

$\dfrac{D_p}{D_B} = N_{Sf}$ is a flow line interception parameter.

In the above, k_m is the Stokes-Cunningham correction factor for mean free path of gas molecules - a factor allowing for correction of slip flow effects around small dust particles;

ρ_p = particle density

D_p = particle diameter

U_o = superficial filter medium velocity

D_B = significant dimension of collecting body

A typical correlation curve for a fiber mat filter corresponding to Equation 1 is shown in Figure 25 in which the target efficiency, which is a function of the overall efficiency, is plotted against N_{Si} with N_{Sf} as parameter, and the reversal trend caused by changes in controlling mechanism is displayed. No correlation of the above type for granular filters has been reported. Since the mechanism of collection on granules is the same as in fiber beds, however, it would appear that a correlation as is shown in Figure 25 could possibly be transposed to a granular bed situation by simply evaluating the dimensionless groups for any granular filter under consideration. This is certainly a reasonable approach, in principle, but there is a great problem with regard to the granule diameter. Note first that the efficiencies plotted in Figure 25 are a strong function of the diameter of the individual collecting fibers. Secondly, the fiber sizes used are considerably smaller than granular particle sizes of greatest potential interest (about 6-60 mesh) from the standpoint of pressure drop. Extrapolation of collection efficiencies found on fiber beds to granular devices using 1/16 inch as the size parameter would give predictions of very poor performance on the latter.* On the other hand, it is not clear that the controlling size dimension for a granular filter should indeed be the granule diameter. Dust collection may be much more influenced by small protruberences or angular sections on the surface of a granule than by the overall dimension. This is brought out since the limited data that are available on granular filtration (Section I-B) indicate that good collection efficiencies can be obtained experimentally and would not be predicted from extrapolation of data on fiber beds to larger diameter collectors.

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* For the fiber system of Figure 25, a value of $\eta_t = 0.01$ corresponds to an overall collection efficiency of 32% for a 10 μ fiber, and a value of $\eta_t = 0.10$, to an efficiency of 98.1%.

B. Derivation of Sorption Models

1. Counterflow Gas-Solid Contactor

The situation to be modeled is shown in Figure 26. Solid sorbent, completely regenerated and flowing at mass velocity L lbs/hr-ft^2, enters the sorption section at the top. The distance variable is measured in the direction of sorbent flow. The dilute gas to be treated flows from the bottom up through the sorption section at mass velocity G.

The governing SO_2 balances are

Gas side $$\frac{dy}{dx} = \left(\frac{\rho_B}{G}\right)\frac{dW}{dt} \qquad (1)$$

$$\frac{dW}{dx} = \left(\frac{\rho_B}{L}\right)\frac{dW}{dt} \qquad (2)$$

The governing rate expression is

$$\frac{2.14}{y}\frac{dW}{dt} = \left(\frac{675}{1 + 6.2S}\right)(1 - S^2) \qquad (3)$$

In the above equations, the following notation has been used:

y = mass fraction SO_2

x = distance measured from solid inlet, ft

ρ_B = bulk density of sorbent, lb/ft^3

G = mass velocity of gas, lb/hr ft^2

W = sorbent loading of SO_2, lb SO_2/lb sorbent

W_o = saturation loading of sorbent, lb SO_2/lb sorbent

t = time, hrs

L = mass velocity of solid, lb/hr ft^2

$S = W/W_o$, fractional saturation of sorbent

An overall material balance yields:

$$LW_2 = G\,(y_2 - y_1) = G\Delta y \text{ or}$$

$$LS_2\,W_o = G\Delta y \qquad (4)$$

The differential equations can be put into dimensionless form by the introduction of a scaling factor $\eta = x/z$, where z is the length of the worbent section in feet.

Combination of equations 2, 3, and 4 gives

$$ds/d\eta = (B)\ (AS + 1)\ \ (1 - S^2)\ \ (675/1 + 6.2S) \qquad (5)$$

where $A = W_oL/y_1G$ and $B = \rho_B y_1 Z/2.14W_oL$

Integration of equation (5) leads to

$$\eta = \frac{1}{675B}\left[\frac{N}{A}\ln(AS + 1) - M\ln(1 - S) + P\ln(1 + S)\right] \qquad (6)$$

In equation 6

$N = (A)\ (A - 6.2)\ /(A^2 - 1)$

$M = 3.6/A + 1$

$P = 2.6/A - 1$

Note that since the factor Z appears in the denominator of both sides of the equation (6) (in η and B, respectively), the length of the sorption path required for a given service does not depend on the overall packed length of the apparatus.

For the calculation of Section III, the method of approach is straightforward. The values used were

$L = 5.72$ lbs/hr ft^2

$G = 111$ lbs/hr ft^2

$W_o = 0.32$ lbs SO_2/lb sorbent

$Z = 0.341$ feet

$y_1 = 4.26 \times 10^{-4}$ lbs SO_2/lb gas

$y_2 = 4.26 \times 10^{-3}$ lbs SO_2/lbs gas

$\rho_B = 45$ lb/ft^3

By calculation

$A = 38.8$ $B = 0.00167$ $N = 0.840$

$M = 0.0905$ $P = 0.0688$

Applying the material balance equation (4) gives $S_2 = 0.233$ and then from (6), $x = 0.0266$ feet.

2. Crossflow Contactor

The situation to be modeled is shown in Figure 27. Solid sorbent completely regenerated and flowing at mass velocity L lbs/hr-ft^2 enters the sorption section at the top flowing through areas A_s. The gas flows from left to right through the shaft of area A_G. The two distance coordinates have their origin in the upper left-hand corner and are positive downward and to the right, respectively. Although the SO_2 content of both phases is uniform where they respectively enter the sorption zone, both will display concentration profiles at their exit sections.

The governing differential SO_2 balances are

$$\frac{\partial y}{\partial x} = \frac{-\rho_B}{G}\frac{\partial W}{\partial t} \text{ - gas side} \qquad (8)$$

$$\frac{\partial W}{\partial z} = \frac{\rho_B}{L}\frac{\partial W}{\partial t} \text{ - solid side} \qquad (9)$$

with boundary conditions

at $x = 0$, $y = y_o$

at $z = 0$, $W = 0$

The overall material balance is

$$LA_s \bar{W}_2 = GA_G (y_o - \bar{y}_2) \qquad (10)$$

where $\bar{W}_2$ and $\bar{y}_2$ are the average outlet SO_2 concentrations in the solid and gas phases respectively (other notation as before).

The rate expression is again

$$\frac{1}{y}\frac{\partial W}{\partial t} = \left(\frac{a}{1 + bS}\right)(1 - S^2) \qquad (11)$$

where $a = 675/2.14 = 315$ and $b = 6.2$

Equations 8 and 9 with their boundary conditions can be put into dimensionless form to yield

$$\frac{\partial U}{\partial \eta} = -\alpha U (1 - S^2)(a/1 + bS) \qquad (12)$$

and

$$\frac{\partial S}{\partial \tau} = \gamma U (1-S^2)(a/1+bS) \qquad (13)$$

with

$U = 1$ at $\eta = 0$ (A)

and

$S = 0$ at $\tau = 0$ (B)

In the above equations,

$$U = \frac{y}{y_o}, \quad \eta = \frac{x}{l}, \quad \tau = \frac{z}{h}$$

$$\alpha = \frac{l \rho_B}{G} \quad \text{and} \quad \gamma = \frac{h \rho_B y_o}{W_o l}$$

The solution is effected as follows: Equation 13 is solved for ∂U. Both sides are differentiated with respect to η. Then noting from (12) and (13) that

$$\frac{1}{\alpha} \frac{\partial S}{\partial \tau} = - \frac{1}{\gamma} \frac{\partial U}{\partial \eta}$$

we eliminate $\partial U/\partial \eta$. The result is a perfect differential that is integrated first in τ and then in η using boundary condition (B). This yields

$$\ln S + (b-1/2) \ln (1 + S) - (b+1/2) \ln (1 - S) + a\alpha\eta =$$

$$\ln S_{(0,\tau)} + (b-1/2) \ln (1 + S_{0,\tau}) - (b+1/2) \ln (1-S_{0,\tau}) \quad (14)$$

$S_{(0,\tau)}$ is the value of S at the boundary of the apparatus where the gas enters, i.e., at x = 0.

Next, returning to equation (12) and applying boundary condition (A), the result is integrated with respect to τ leading to

$$2a\gamma\tau = (1-b) \ln (1+S_{0,\tau}) - (1+b) \ln (1-S_{0,\tau}) \quad (15)$$

which is an implicit expression of $S_{0,\tau}$. Equations (14) and (15) can then be solved for S as a function of τ and η. These equations were applied to a sample calculation for a 9' x 14' x 1/2' crossflow cell operated at a gas velocity of 1.5 ft/sec, with an entering SO_2 mass fraction of 4.26×10^{-3}. The gas density at 300°C was taken as 0.0398 lbs/ft^3. The sorbent, alkalized alumina, was again taken to have a bulk density of 45 lb/ft^3 and an SO_2 capacity of 0.32 lb SO_2/lb sorbent. For a 200 minute granule residence time in the bed, a solid flow rate of 189 lb/hr ft^2 was calculated. Using these values $S_{0,\tau} = 0.982$ at the bottom of the bed where $\tau = 1.0$ by equation (15). The next step was to calculate the outlet sorption profile from equation (14) at $\tau = 1$ for various values of η. The results of this calculation are shown in Figure 28; the sorption front has not penetrated the full thickness of the bed under these conditons. Graphical integration of Figure 28 then gave a value of S_2 of 0.425, and thus, W_2, of 0.136. Substituting the appropriate values into equation (10), the overall material balance, gives a value of $\bar{y}_2$, the average outlet gas concentration, very close to zero indicating a near complete removal of SO_2.

3. Fixed-Bed Contactor

The situation to be modeled is shown in Figure 29. A flow of gas at G lbs/hr ft^2 and entering SO_2 mass fraction y_o enters a stationary bed of regenerated sorbent of height Z.

The governing equations in this case are

$$\frac{\partial y}{\partial x} = -\frac{\rho_B}{G}\frac{\partial W}{\partial t} \quad \text{- gas side } SO_2 \text{ balance} \tag{16}$$

$$\frac{\partial W}{\partial t} = y\ (1-S^2)\ (a/1+bS) \tag{17}$$

with

$$y = y_o \text{ at } x = 0$$

and

$$S = 0 \text{ at } t = 0$$

Equating the total sorbent capacity with total gas loading, we obtain

$$Z\ \rho_B\ W_o = G\ y_o\ \theta_B$$

or

$$\theta_B = Z\ \rho_B\ W_o/G\ y_o$$

The quantity θ_B is the time at which the total bed capacity for SO_2 is reached assuming complete sorption of all contacting SO_2. It is not a breakthrough time and does not indicate the time for a given outlet concentration of SO_2 to be exceeded.

The differential equations can be put into non-dimensional form by the substitution of

$$\tau = \frac{t}{\theta_B}, \quad U = \frac{y}{y_o}, \text{ and } \eta = \frac{x}{z}$$

and defining $\alpha \equiv \frac{Z\ \rho_B}{G}$

We thus obtain

$$\frac{\partial S}{\partial \tau} = \alpha\ U\ \ (1-S^2)\left(\frac{a}{1+bS}\right) \tag{18}$$

and

$$\frac{\partial U}{\partial \eta} = -\alpha\ \ U\ (1-S^2)\ \left(\frac{a}{1+bS}\right) \tag{19}$$

with

$$U = 1 \quad \text{at } \eta = 0 \quad \text{at any } \tau \qquad \text{(C)}$$

and

$$S = 0 \text{ at } \tau = 0 \quad \text{at any } \eta \qquad \text{(D)}$$

The solution is obtained as follows: Equation (18) is differentiated with respect to η and $\partial U/\partial\eta$ is eliminated through substitution of (19). The resulting equation is

$$\frac{\partial}{\partial\tau}\left[\left(\frac{1}{1-S^2}\right)\left(\frac{1+bS}{a}\right)\frac{\partial S}{\partial\eta} + \alpha S\right] = 0 \qquad (20)$$

Equation (20) is then integrated, first with respect to τ and applying boundary condition (D), and then in η to give

$$\begin{aligned} &\ln S + \left(\frac{b-1}{2}\right)\ln(1+S) - \left(\frac{b+1}{2}\right)\ln(1-S) + a\alpha\eta \\ &= \ln S(0,\tau) + \left(\frac{b+1}{2}\right)\ln\left(1+S_{(0,\tau)}\right) - \left(\frac{b+1}{2}\right)\ln\left(1-S_{(0,\tau)}\right) \end{aligned} \qquad (21)$$

or

$$\ln S + \left(\frac{b-1}{2}\right)\ln(1+S) - \left(\frac{b+1}{2}\right)\ln(1-S) + a\alpha\eta = C(\tau) \qquad (22)$$

where

$$C(\tau) \equiv \ln S_{(0,\tau)} + \left(\frac{b-1}{2}\right)\ln\left(1+S_{(0,\tau)}\right) - \left(\frac{b+1}{2}\right)\ln\left(1-S_{(0,\tau)}\right) \qquad (23)$$

In order to calculate gas phase behavior, differentiate (23) with respect to τ and combine with (18) to obtain

$$U = \frac{(1+bS)}{a\alpha(1-S^2)}\frac{\partial C(\tau)}{\partial\tau}\left[\frac{1}{\frac{1}{S} + \left(\frac{b-1}{2}\right)\left(\frac{1}{1+S}\right) + \left(\frac{b+1}{2}\right)\left(\frac{1}{1-S}\right)}\right] \qquad (24)$$

which reduces to

$$U = (1/a\alpha)\,(\partial C/\partial\tau)\,(S) \qquad (25)$$

In order to calculate $C(\tau)$ or $\partial C(\tau)/\partial\tau$, $S(0,\tau)$ must be evaluated. This is handled by returning to equation (18), setting $\eta = 0$ to get

$$\alpha\, U(\eta=0) = \left(\frac{1}{1-S^2_{\ (\eta=0)}}\right)\left(\frac{1+bS_{(y=0)}}{a}\right)\left(\frac{\partial S_{(\eta=0)}}{\partial\tau}\right) \qquad (26)$$

From boundary condition (c), since $U = 1$ when $\eta = 0$

$$\alpha = \left(\frac{}{1-S^2_{\ (\eta=0)}}\right)\left(\frac{1+bS_{(\eta=0)}}{a}\right)\left(\frac{\partial S_{(\eta=0)}}{\partial\tau}\right) \qquad (27)$$

Integrating (27) from 0 to τ we obtain

$$2a\alpha\tau = (1-b)\ln\left(1 + S_{(0,\tau)}\right) - (1+b)\ln\left(1 + S_{(0,\tau)}\right) \tag{28}$$

The calculation procedure is as follows

1. Determine values of τ corresponding to inserted values of $S_{(0,\tau)}$ in equation (28). Obtain curve of $S_{(0,\tau)}$ vs τ.
2. From the above curve, calculate $C_{(\tau)}$ from equation (23).
3. This enables a calculation of S at any η for particular values of τ using equation (22).
4. From a plot of $C_{(\tau)}$ vs τ, the value of $\partial C_{(\tau)}/\partial\tau$ can be obtained for any value of τ. This is used in equation (25) in conjunction with S values obtained from equation (22) (Step 3) in order to calculate U at any point in the bed at any time.

To determine the breakthrough concentrations at the bed outlet as a function of time:

Carry out the above procedure, but set $\eta = 1$ in equation (22).

A calculation was carried out for a 1/2 foot thick bed with a gas velocity of 1.5 ft/sec. Again using values of

ρ_B = 45 lbs/ft^3

ρ_G = 0.0398 lbs/ft^3

y_0 = 4.26×10^{-3} lbs SO_2/lb gas

a = 315 and b = 6.2

G = 215 lb/hr ft^2

θ_B = 7.85 hrs

α = 0.1047

If $S_{(0,\tau)}$ is taken as 0.999, a value of $\tau = 0.7$ results, i.e., the bed has been run for (7.85) x (0.7) = 5.5 hours. From equation (23), $C_\tau = 26.7$. Substituting this value into equation (22) and setting $\eta = 1$ leads to a value of S at the top of the bed of 0.0017. Thus, even after 5.5 hours, the center of the sorption front has not penetrated to the top of the 6" bed and the sorbent at that point is less than 0.2% saturated.

Finally, a plot of $C_{(\tau)}$ vs τ gives a value of $\partial C_{(\tau)}/\partial\tau = 33$, very close to the value for $(a\alpha)$, and thus by (25), U at the bed outlet is also 0.0017. In other words, less than 0.2% of the SO_2 in the gas is leaving the sorber after 5.5 hours of operation.

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TABLES

Table I

Conditions for 800 mw Power Station Gas Flow — 87.3 Million Standard Cubic Feet per Hour (3)

Gas Temperature — 300°C

Gas Composition — Volume %

N_2	76.2
O_2	3.4
CO_2	14.2
H_2O	6.0
SC_2	0.2
SO_3	Trace

Flyash loading — 3.3 grain/std ft^3 (13)

Approximate flyash size distribution (16)

Wt. %	Particle Diameter, Microns
9.5	$<$ 2.0
8.5	2.0 — 3.4
11.0	3.4 — 6.3
10.0	6.3 — 11.8
9.0	11.8 — 17.0
6.0	17.0 — 24.0
19.0	24.0 — 43.2
18.0	43.2 — 73.0
6.1	73.0 — 150.0
2.9	$>$ 150

Table II

RESULTS OF HANFORD SAND FILTER TESTS (14)

Effect of Depth of Sand

Depth of Sand — Feet

Depth of Sand — Feet	Collection Efficiency, % (Laboratory Beta Analysis)
1	91.6
2	99.5
3	99.8
4	99.9

Effect of Sand Size

Sand Type	Average Diameter, Inches	Superficial Gas Velocity, Ft./min.	Collection Efficiency, % (Laboratory Beta Analysis)
Hanford 16-20 mesh	0.040	3.8	96.3
Hanford 20-40 mesh	0.022	3.8	99.98
Hanford 16-20 mesh	0.040	10.0	89.5
Hanford 20-40 mesh	0.022	10.0	99.88
Ottawa 20-30 mesh	0.028	5.1	96.2
Ottawa 30-40 mesh	0.019	5.1	98.6
Ottawa 20-30 mesh	0.028	10.0	92.9
Ottawa 30-40 mesh	0.019	10.0	98.2

Effect of Type of Sand

Type Sand	Collection Efficiency, %
Hanford 20-40 mesh	99.93
A.G.S. Flint 30-40 mesh	99.84
Monterey Type G	99.61
Eau Claire Type G	99.38

Effect of Gas Velocity

Superficial Gas Velocity ft/min	Collection Efficiency, %	
	Laboratory-beta	Instrument
1.8	99.97	99.90
6.1	99.83	99.80
10.0	98.9	99.0
18.3	92.7	97.0
40.0	97.0	96.0

Table III

Value of Coefficient for Hanford Correlation

Type of Sand	K
Hanford	0.053
A. G. S. Flint	0.045
Rounded grains (Ottawa, Eau Claire, Monterey)	0.035

Table IV

Data Obtained from Strauss Granular Filter Study(8)

Results of pilot plant experiments - individual test with bed 1-10½" of crushed high-temperature insulating brick graded in the range 5/16" - 7 B.S.S. mesh

Test No.	Furnace Operation	Test Time Minutes	Mass Rate of Gas Flow, $lb/ft^2/hr$	Gas Velocity ft/sec.	Gas Temperature, °C	Inlet Fume Concentration g/nm^3	Fe_2O_3 in fume wt %	Bed Thickness in.	Filter Bed Efficiency, wt %	Efficiency in Fe_2O_3 deposition, wt %	Max. Bed Resistance in water gauge
3	R*	13.0	226	1.43	230	1.63	----	10½	90.5	----	2.20
4	R	15.0	204	1.20	230	4.14	----	10½	96.3	----	2.19
5	C	19.0	225	1.47	275	0.123	----	10½	87.3	----	2.60
8	O	4.88	172	1.19	290	6.26	----	10½	87.6	----	3.50
9	R	13.13	145	0.96	270	3.83	----	10½	94.7	----	4.00
10	O	6.95	140	0.79	237	6.24	78.0	9	85.0	87.3	4.70
12	R	15.1	214	1.67	345	0.93	64.0	9	90.2	96.0	2.20
14	A	14.4	212	1.54	303	0.54	20.0	9	84.0	97.1	2.40
15	T	12.7	204	1.52	317	0.60	23.0	9	87.5	98.1	2.50
32	M	24.9	126	0.92	305	0.127	36.0	9	84.6	85.7	4.90
36	M	20.4	161	1.26	352	0.163	44.2	9	74.3	74.1	3.60
37	R	18.3	200	1.50	300	1.24	55.2	9	94.2	----	3.90
41	R	20.6	338	3.35	520	1.73	----	3	85.3	----	2.10
45	R	13.0	206	1.29	225	0.595	----	3	65.5	----	0.70
49	F	10.4	317	2.94	460	0.232	----	3	82.7	----	2.35
56	R	15.1	256	2.09	370	0.573	----	1	59.3	----	0.38
71	R	22.2	314	2.51	360	1.18	----	1	87.0	----	0.85
80	R	15.9	354	2.30	455	4.94	----	1	88.2	----	0.52

* These letters refer to the phase of the steelmaking cycle in progress at the time of test.

Table V

Operating Characteristics of an Installation of the Dorfan Impingo Filter

Type of Service — collection of dust generated during the drying of asbestos rock

Shaft Size — 9' x14' filtering area, shaft 1' thick

Capacity — 32,000 cfm

Inlet Dust Loading — 6 grains/std ft^3

Size of Dust — 100% through 100 mesh
60% finer than 10 microns

Collection Granules — raw asbestos rock ranging from 1/2" to 1-1/2" in size

Typical Residence Time of Granules — 200 minutes in shaft

Collection Efficiency — 97%

Temperature and Pressure Drop Information — Not given

Table VI

Typical Operating Characteristics of Consolidation Coal Co., Shaft Filter

Type of Service — Recovery of dust from coal drying operation

Size — Individual cell was 4'2" wide x 4' high x 20" thick.
In the unit the filtering surface was divided into two banks of six (6) cells each in parallel.

Capacity — 100-300 cfm/ft^2 filter area

Total Filter Area — 200 ft^2

Inlet Dust Loading — 3-7 gr/scf

Size of Dust — 86% $<30\mu$ and 22% $<10\mu$

Collection Granules — Coal, 1-1/4"- 3/8"

Granule Flow Rate — 0.035 cfh/scfm gas (approximately 24 tons/hour)

Operating temperature — 130°F

Pressure Drop — 2.3 in wg

Collection efficiency — 89 - 99 + %
(Excluding attrited coal from efficiency calculation)

Table VII

Operating Characteristics of Carnegie-Mellon Filter During SO_2 Sorption

Shaft Size — One foot square with one foot high filtering section

Gas Flows — 60-90 cfm/ft^2

Inlet Dust Loading — 1 grain/scf

Collection Granules — 1/16" - 1/8" alkalized alumina particles mixed with crushed limestone 1/4" - 1/2"

Operating Temperature — 400^oF

Table VIII

Carnegie-Mellon Crossflow Reactor Results

Operating Temperature — 400°F
Gas — Air loaded with SO_2 and flyash
Sorbent — 1/16" - 1/8" alkalized alumina (from USBM)
Filter Area — 1 ft^2

Gas Flow, scfm	31.0	31.0
Solid Rate, lbs/hr	10.5	19.6
Solid Residence Time, Min.	270	144
SO_2 Inlet Concentration, mol%	0.6	0.6
SO_2 Outlet Concentration	Not found	Not found
Inlet Dust Loading, gr/std ft^3	2.82	3.10
Dust Collection Efficiency	99.3%	98.7%
Pressure Drop, in w.g.	2	2

Table IX

Typical Data From Tests On L-S Filter

Application — Removal of flyash from room air

Panel Size — 5' x 1' x 2-3/8" thick

Gas Rate — 12 cfm/ft^2

Inlet Dust Loading — 1.9 $grain/ft^3$

Size of Dust — As per Table I

Collection Granules — 16-30 mesh sand

Puffback Tank Pressure — 40 psig

Pressure Drop — Initial 0.07 in wg, final 0.54 in , after puffback 0.08 in

Collection Efficiency — 99.7%

Table X

Summary of Laboratory Tests on Ducon Filter

Capacity of Unit — 80 - 100 cfm

Filter Velocity — 50 cfm/ft^2 filter area

Pressure Drop — Initial 1.5-2 in w.g., final 4-6 in w.g.

Cycle Time — 3/4 - 1 hour

Material	Inlet Loading gr/SCF	Outlet Loading gr/SCF
Iron Oxide from Oxygen - lanced Electric Furnace	3 - 4	0.003 - 0.03
Iron Oxide from Oxygen-lanced Open Hearth Furnace	2 - 5	0.02 - 0.07
Low Grade Nickel Ore from Aerofall Mill (Approx. 25% $< 5\mu$ 10% $< 2\mu$	5	0.001 - 0.01
Flyash From Coal-fired Boiler	3 - 3.5	0.03 - 0.07
Talc Dust (20% $< 1\mu$)	3 - 3.5	0.02 - 0.05
Plastic (PVC) Dust from Mfg. Process	2 - 3	0.01 - 0.05

Table XI

Typical Operating Characteristics of Lurgi-MB-Filter

Type of Service — Cleaning of waste air from cement clinker coolers

Gas Rate — 45,000 cfm

Inlet Dust Loading — 2.2 - 2.6 gr/ft^3

Collection Granules — About 1/8" coarse gravel particles

Operating Temperature — 480 - 660°F

Pressure Drop — 2 in w.g.

Collection Efficiency — 99.2%

Table XII

Calculations for GSC Sorber

Gas Rate — 15 ft/min or 111 lbs/hr-ft^2

Solid Rate — 5.72 lb/hr ft^2 *

Entering Gas Concentration — 0.2 mol % SO_2

Outlet Gas Concentration — 0.02 mol % SO_2

Entering Solid Concentration — 0 lbs. SO_2/lb sorbent

Outlet Solid Concentration — 0.0746 lb SO_2/lb sorbent

Outlet Solid Fractional Saturation — 0.233

Pressure Drop — 3" w.g.

---

* (Based on an entering gas loading of 3.3 grains/std ft^3, and a ratio of sorbent used to flyash collected of 4 to 1 as per the estimate of A. M. Squires.)

Table XIII

Calculation Summary for Crossflow Model

Dimensions of Unit — 9' x 14' x 1/2'

Gas Rate — 1-1/2 ft/sec or 215 #/hr ft^2

Solid Rate — 189 #/hr ft^2 (based on 200 minute residence time)

Entering Gas Concentration — 0.2 mol % SO_2

Outlet Gas Concentration — None

Entering Solids Concentration — None

Outlet Solid Concentration — 0.137 lb SO_2/lb sorbent

Outlet Solid Fractional Saturation — 0.428

ΔP — 7" w.g.

Table XIV

Fixed Bed Calculation Summary

Bed Thickness — 6"

Gas Velocity — 1.5 ft/sec

Entering Gas Concentration — 0.2 mol % SO_2

Outlet Gas Concentration — 0.0034 mol %

Time of Operation — 5.5 hours

Table XV

Capital Costs — Squires Sorbers and Dust Collectors

Basis: 800 mw power station
3×10^6 cfm gas
2.46×10^6 tons coal per year
Filtering face velocity — 15 ft/min
Puffback cleaning system
300°C

	Sorber-Collector		Dust Collectors - New Plant	
	Existing Power Station	New Power Station	GSC	LS
Panels	$1,823,500	$1,823,500	$1,823,500	$1,192,000
Enclosures	36,700	36,700	36,700	28,800
Internal Puffback Ductwork	194,000	194,000	194,000	194,000
External Ducting	300,000	300,000	300,000	300,000
Dampers	270,000	270,000	270,000	270,000
Compressors and Motors	58,500	58,500	46,500	46,500
Cyclones and Solid Flow Valves	20,500	20,500	2,100	2,100
Screw Conveyors	4,900	4,900	4,900	4,900
Sorbent Storage Tank	2,500	2,500	2,500	2,500
Flyash Collection Tank	37,200	37,200	37,200	37,200
Puffback Dampers	45,000	45,000	45,000	45,000
Power Plant Blowers	12,000	33,000	33,000	17,000
Subtotal	2,804,800	2,825, 00	2,795,400	2,140,000
Subtotal After Factor Applied	8,694,900	8,760,000	8,665,800	6,634,000
Regeneration Equipment	557,000	557,000		
Gas Producer	645,000	645,000		
Claus Unit	1,780,000	1,780,000		
Total Capital Investment	11,676,900	11,742,000	8,665,800	6,634,000
Per kw	$14.60	$14.69	$10.82	$ 8.30
Per cfm	$ 3.89	$ 3.92	$ 2.89	$ 2.21
Total Space Requirement, ft^2	15,200	15,200	15,200	13,100
cfm per ft^2 plot area	198	198	198	229

TABLE XVI

Operating Costs — Squires Sorbers and Dust Collectors

	Sorber		Dust Collectors - New Plant	
	Existing Power Station	New Power Station	GSC	LS
Capital Charges at 14%	$1,635,000	$1,645,000	$1,215,000	$ 929,000
Labor, Maintenance, Overhead	836,900	836,900	182,000	182,000
Raw Materials	712,000	712,000	27,000	27,000
Utilities	170,130	170,130	64,200	29,200
Total Operating Cost	3,354,030	3,364,030	1,488,200	1,167,200
Per Ton Coal (90% Operating Factor)	$1.47	$1.47	$0.67	$0.52

Table XVII

Capital Costs -- Crossflow Sorbers and Dust Collectors

Basis: 800 mw power station
2.92×10^6 cfm gas
2.46×10^6 tons coal per year
90 ft/min filtering face velocity

	Sorber Existing Power Station	Sorber New Power Station	Dust Collector New Power Station
Panels	$ 382,500	$ 382,500	$ 382,500
Enclosures	28,600	28,600	28,600
External Ducting	346,000	346,000	346,000
Dampers	310,000	310,000	310,000
Compressors and Motors	30,000	30,000	16,800
Cyclones and Solid Flow Valves	23,000	23,000	2,100
Screw Conveyors	36,800	36,800	36,800
Sorbent Storage Tank	3,700	3,700	3,700
Flyash Collection Tank	37,200	37,200	37,200
Power Plant Blowers	304,000	87,600	87,600
Subtotal	1,501,800	1,285,400	1,251,300
Subtotal After Factor Applied	4,654,400	3,984,800	3,879,100
Regeneration Equipment	557,000	557,000	
Gas Producer	645,000	645,000	
Claus Unit	1,780,000	1,780,000	
Total Capital Investment	7,636,400	6,966,800	3,879,100
Per kw	$9.55	$8.70	$4.85
Per cfm	$2.62	$2.39	$1.33
Total Space Requirement ft^2	14,600	14,600	14,600
cfm per ft^2 Plot Area	198	198	198

Table XVIII

Operating Costs -- Crossflow Sorbers and Dust Collectors

	Sorber Existing Power Station	Sorber New Power Station	Dust Collector New Power Station
Capital Charges at 14%	$1,169,000	$ 976,000	$ 543,000
Labor, Maintenance, Overhead	836,900	836,900	182,000
Raw Materials	825,000	825,000	39,000
Utilities	195,600	195,600	103,000
Total Operating Cost	3,026,500	2,833,500	867,000
Per ton coal (90% operating factor)	$1.36	$1.28	$0.38

Table XIX

Comparative Cost Estimates

Basis: 800 mw Power Station, 3 x 10^6 cfm Gas, 2.46 x 10^6 Tons Coal per Year

Sorber-Dust Collectors	Capital Investment per kw	Capital Investment per cfm	Annual Operating Cost 90% Load Factor per Ton Coal
Crossflow — New Station	$ 8.70	$ 2.39	$ 1.28
Crossflow — Existing Station	9.55	2.62	1.36
Squires GSC — New Station	14.69	3.92	1.47
Squires GSC — Existing Station	14.60	3.89	1.47
Squires GSC — 1.0 ft/sec Face Velocity, New Station	7.09	1.89	1.14
Squires GSC Sorber — 2.2 ft/sec New Station	7.19	1.92	1.35
Ducon — Fixed Bed, New Station	23.60	6.20	1.78
Dust Collectors -- New Installations			
Crossflow	4.85	1.33	0.38
Electrostatic ($0.60/cfm)	7.14	1.90	0.45
Electrostatic ($0.70/cfm)	8.30	2.21	0.51
Squires LS	8.30	2.21	0.52
Ducon	10.35	2.68	0.64
Squires GSC	10.82	2.89	0.67

Table XX

Cost Estimates for Stack Gas Sulfur Removal Processes (19)

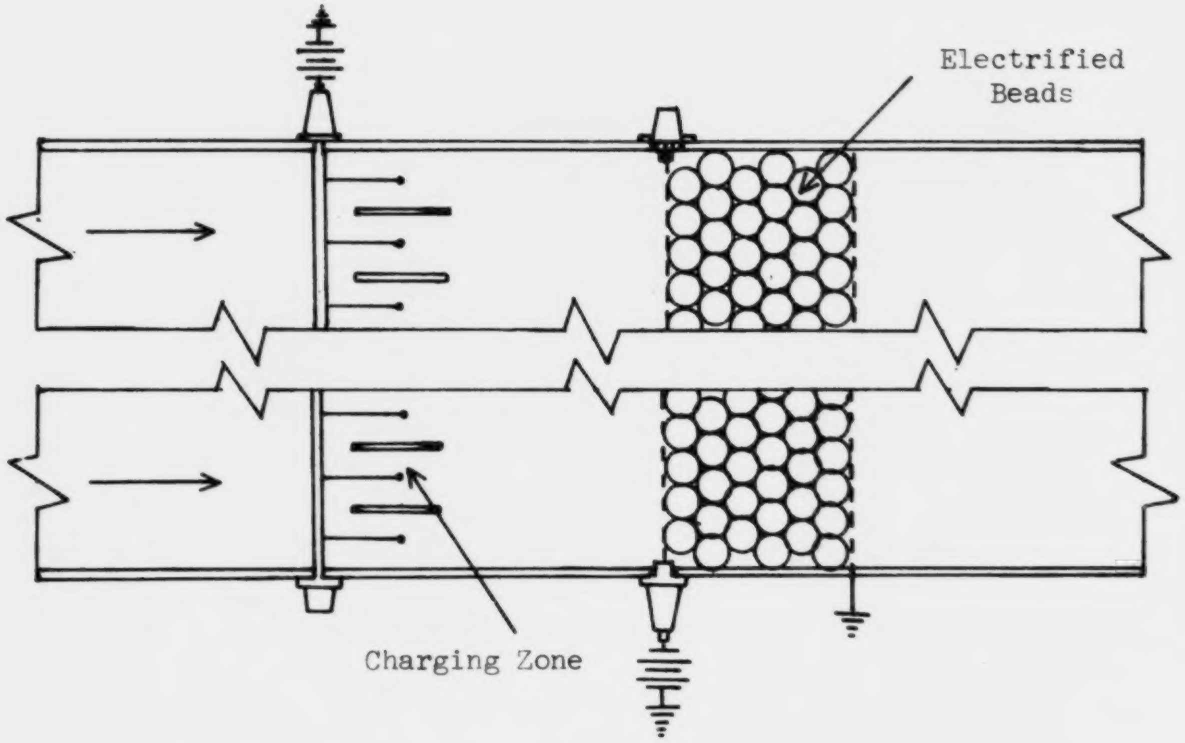
Process	Capital Cost per kw	Operating Cost* per ton
Reinluft	$17.77	$ 2.45
Monsanto	21.25	1.75
BU Mines Alk. Alumina	10.64	1.54
S&W/Ionics	15.40	Not Calculated
Beckwell	8.14	Not Calculated

* 90% Operating Factor

FIGURES

Figure 1.

(from Ref. 2)


Electrified
Beads
Charging Zone

Electrified Packed Bed

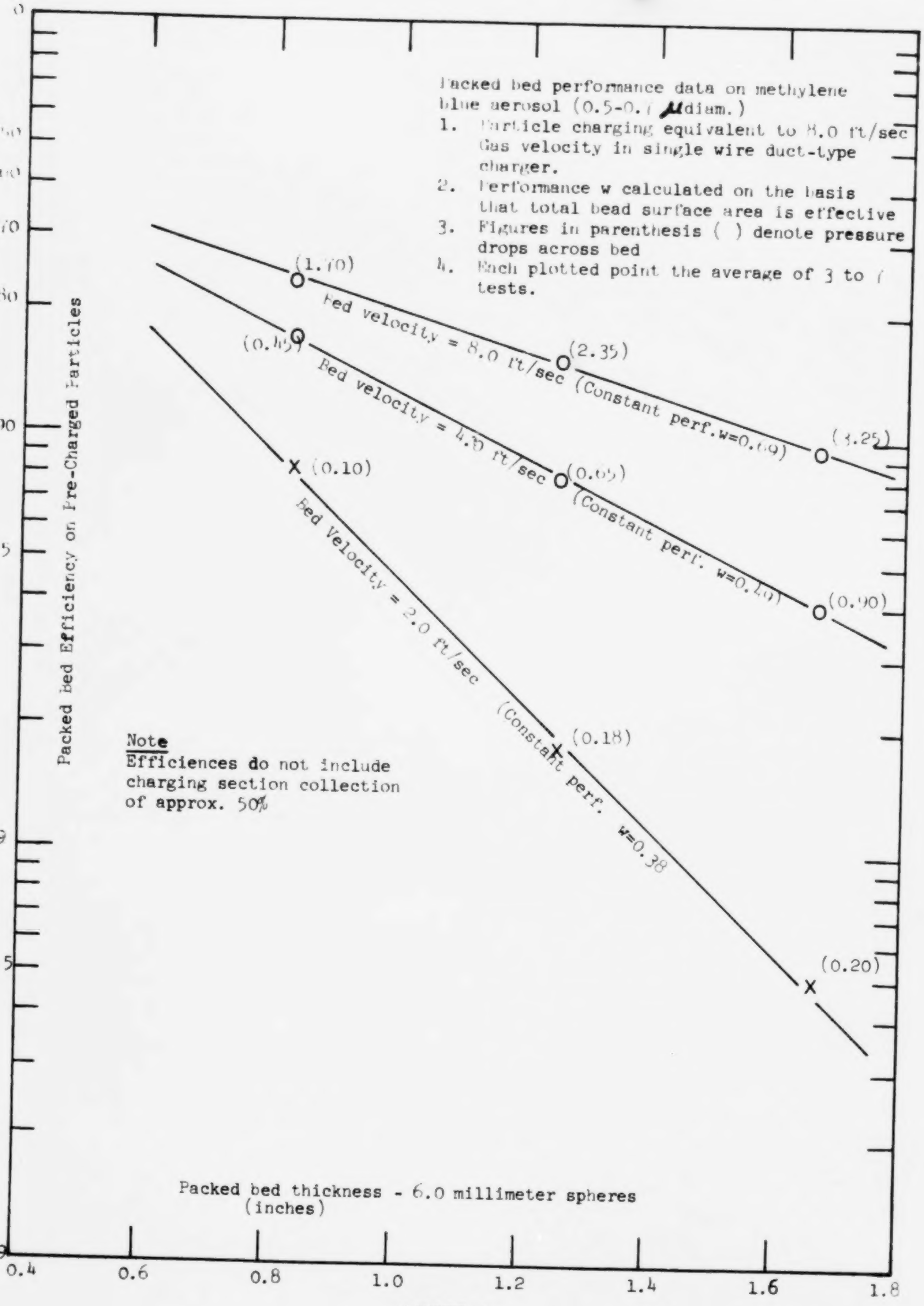

Packed bed performance data on methylene blue aerosol (0.5-0.7 μ diam.)
1. Particle charging equivalent to 8.0 ft/sec Gas velocity in single wire duct-type charger.
2. Performance w calculated on the basis that total bead surface area is effective
3. Figures in parenthesis ( ) denote pressure drops across bed
4. Each plotted point the average of 3 to 7 tests.
Packed Bed Efficiency on Pre-Charged Particles
0
50
60
70
80
90
99.5
99
99.5
99.9
(1.70)
Bed velocity = 8.0 ft/sec (Constant perf. w=0.69)
(2.35)
(3.25)
(0.45)
Bed velocity = 4.0 ft/sec
(0.65)
(Constant perf. w=0.49)
(0.90)
(0.10)
Bed Velocity = 2.0 ft/sec
(Constant perf. w=0.38
(0.18)
(0.20)
Note
Efficiences do not include charging section collection of approx. 50%
Packed bed thickness - 6.0 millimeter spheres (inches)
0.4
0.6
0.8
1.0
1.2
1.4
1.6
1.8

FIGURE 2

FIGURE 3 - PANEL FILTER SECTION

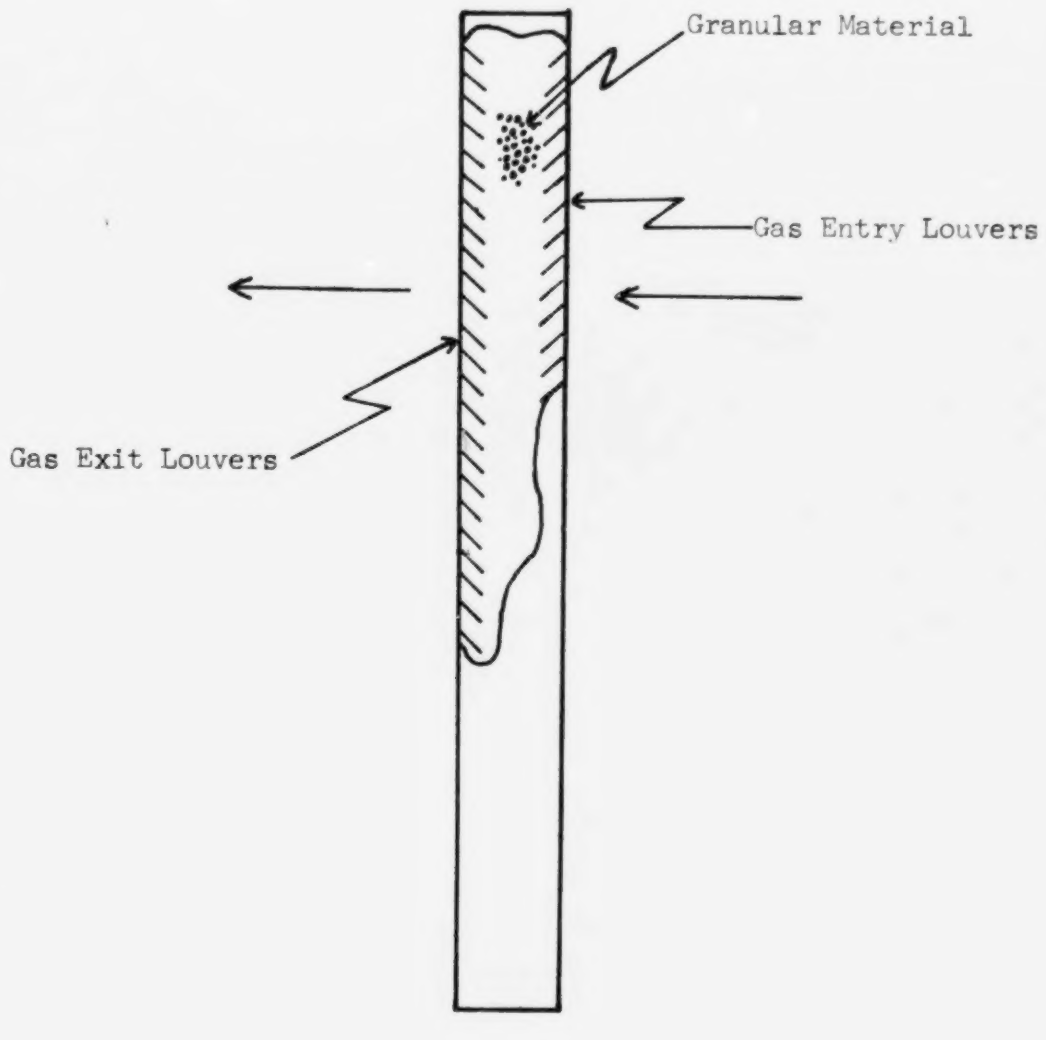

Granular Material
Gas Entry Louvers
Gas Exit Louvers

FIGURE 4

Dorfan Impingo Filter

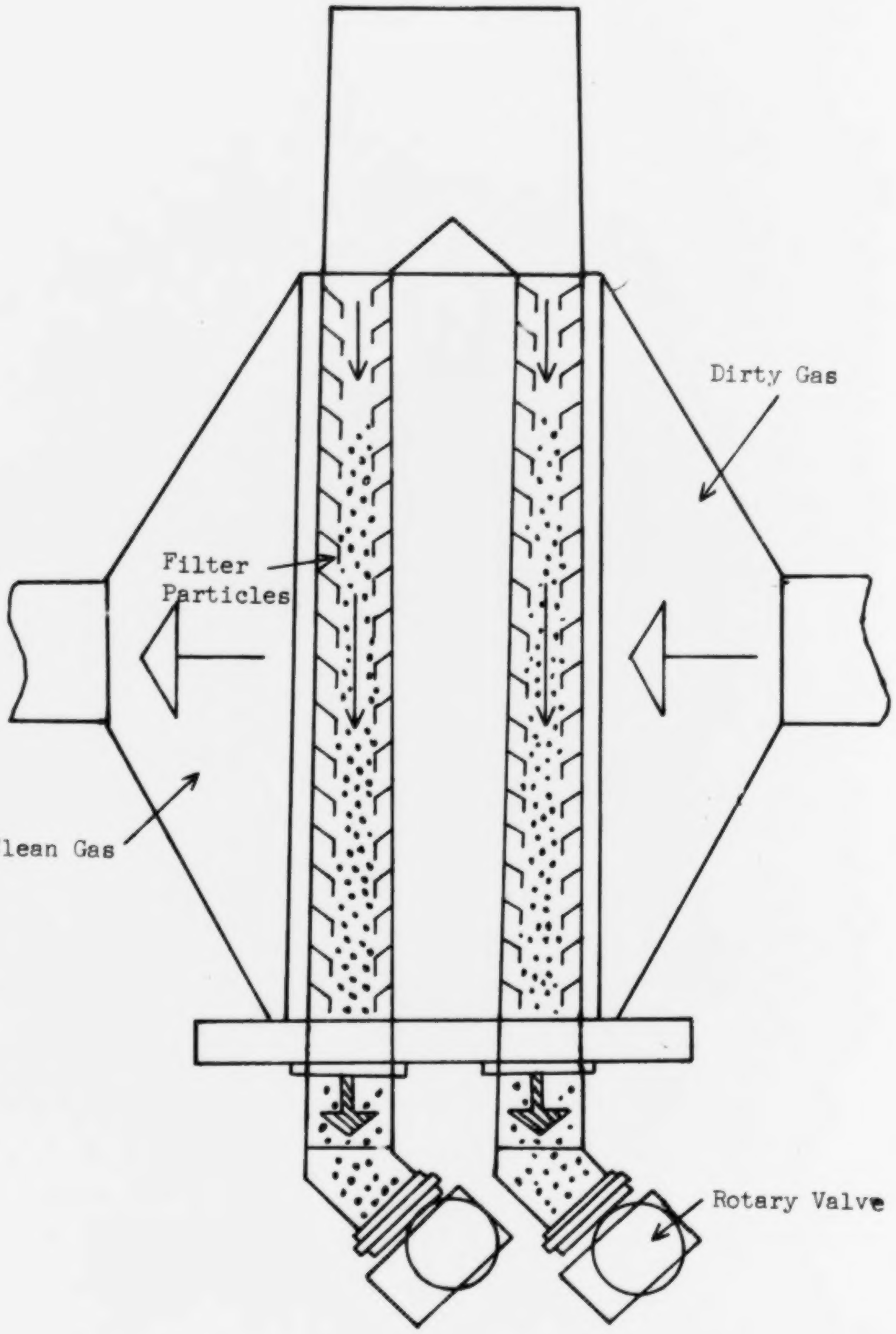

Dirty Gas
Filter Particles
Clean Gas
Rotary Valve

FIGURE 5

Consolidation Coal Company Filter

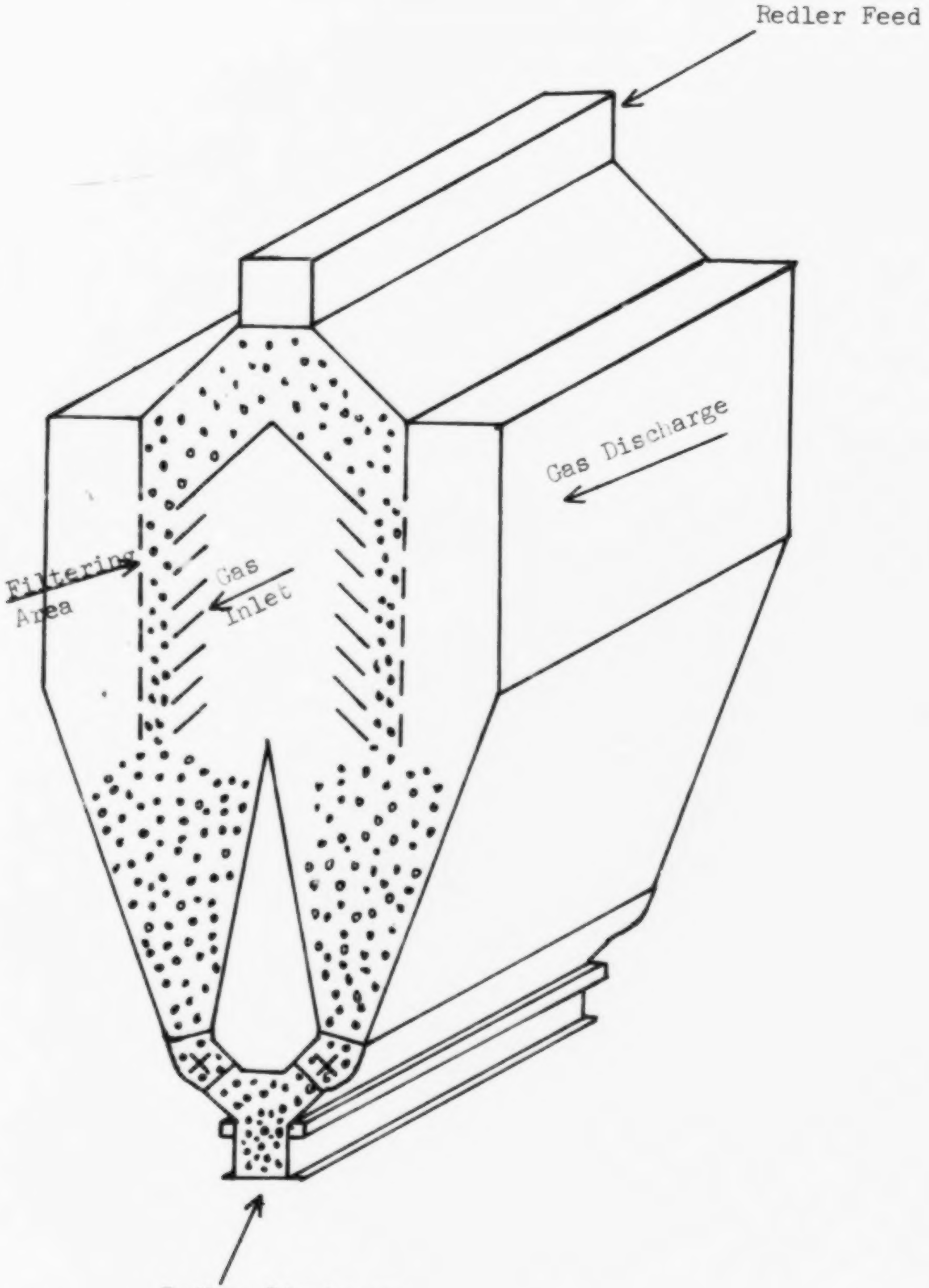
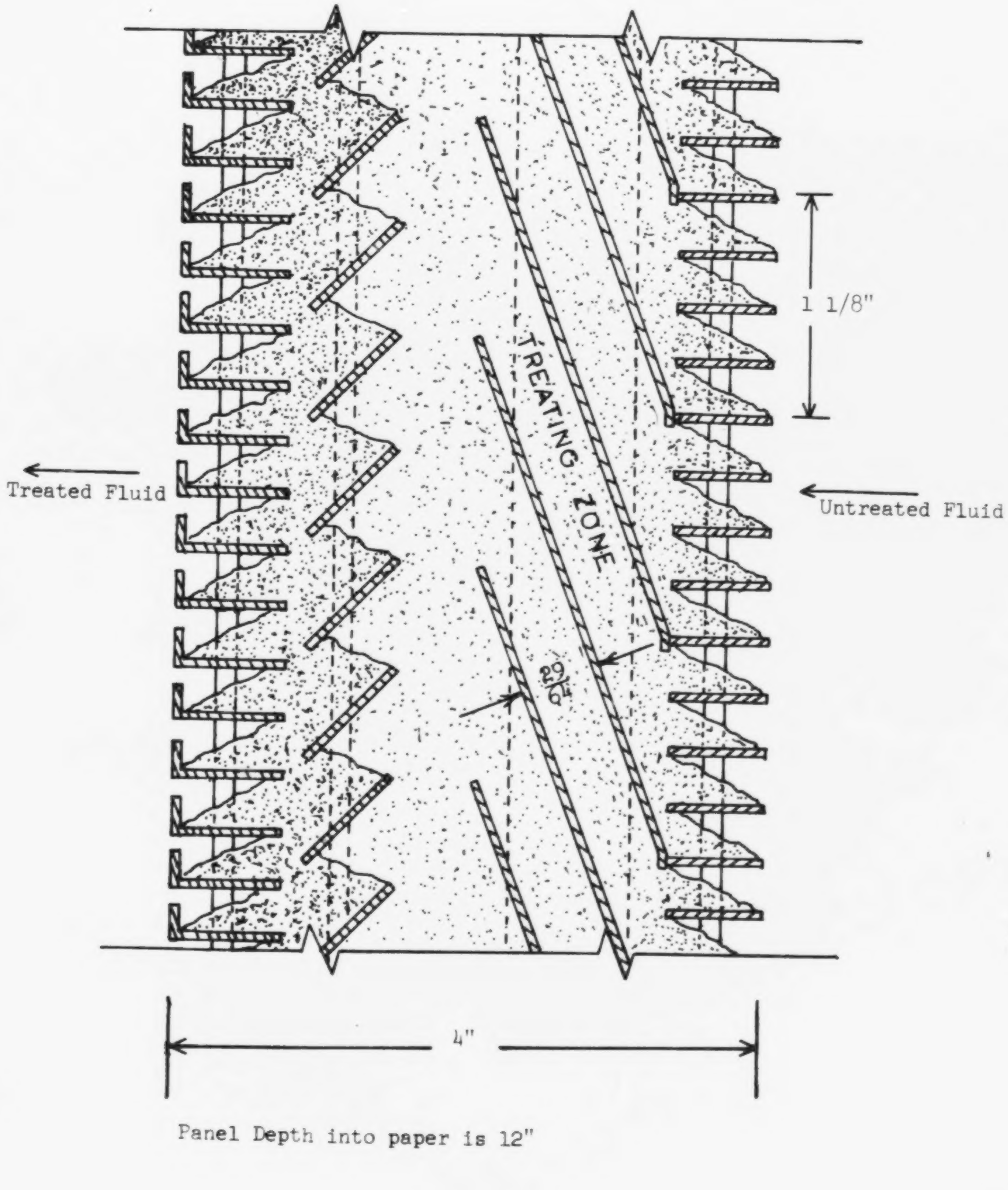

Redler Feed
Gas Discharge
Filtering Area
Gas Inlet
Redler Discharge

Figure 6 Section of the GSC Panel
(Drawn from Patent Application of A.M. Squires)


TREATING ZONE
1 1/8"
Treated Fluid
Untreated Fluid
29/64
4"

Panel Depth into paper is 12"

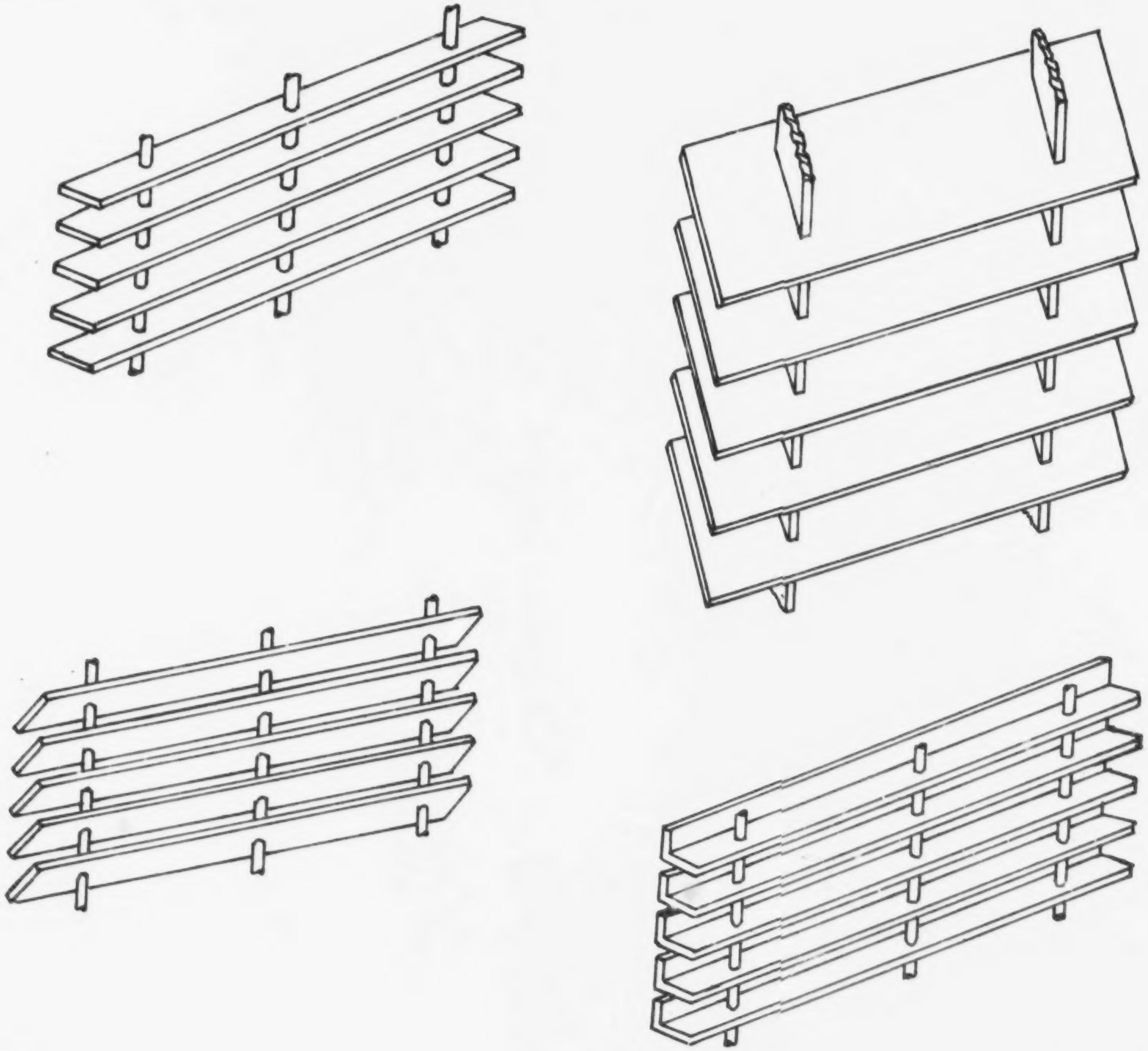


Figure 7 Internal Support Elements for GSC Panel
(From Patent Application of A.M. Squires)

Figure 8.

Arrangement of Two GSC Panels Equipped with Mechanical Means for Bed Cleaning

(From Patent Application of A.M. Squires)

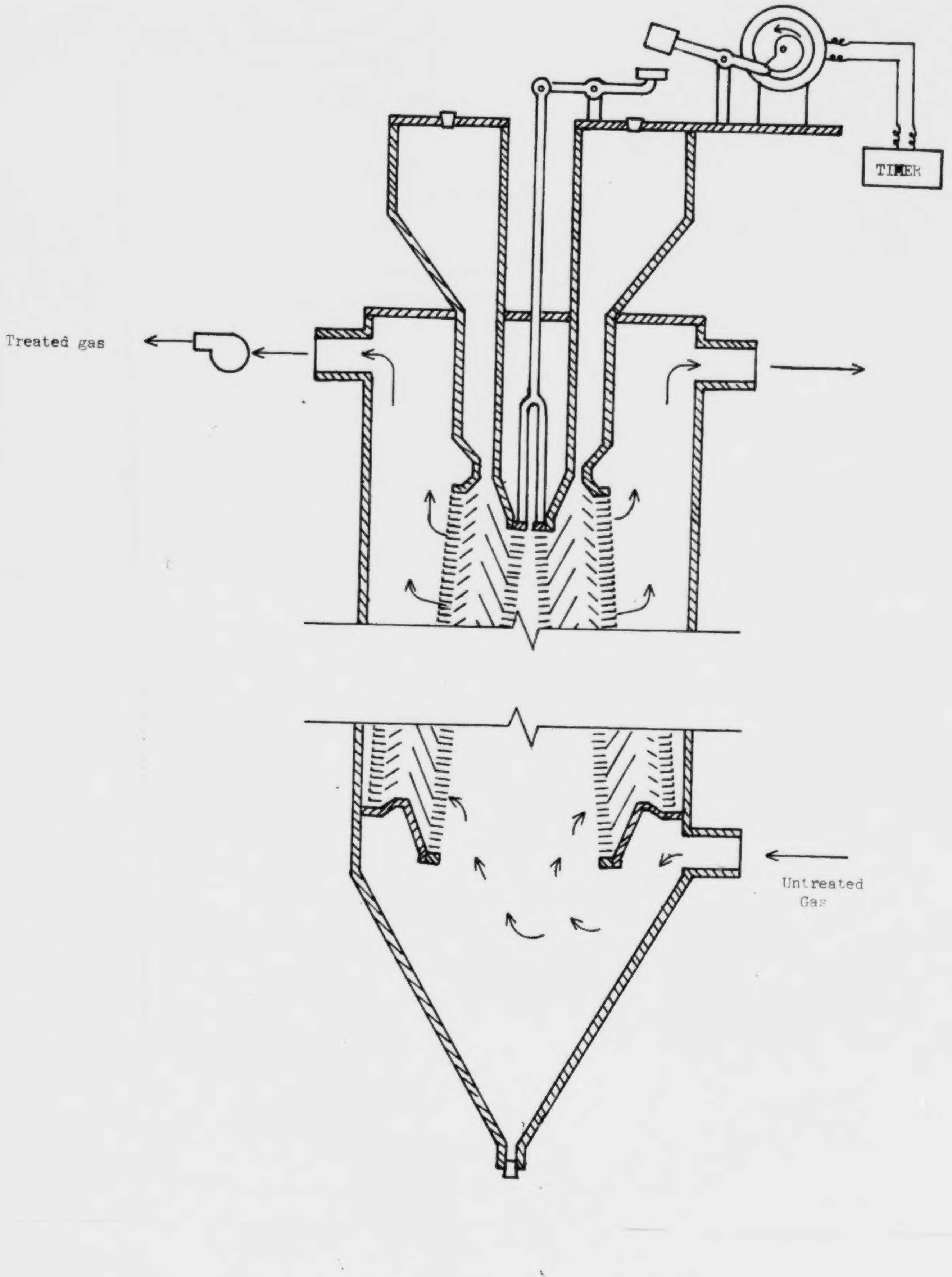
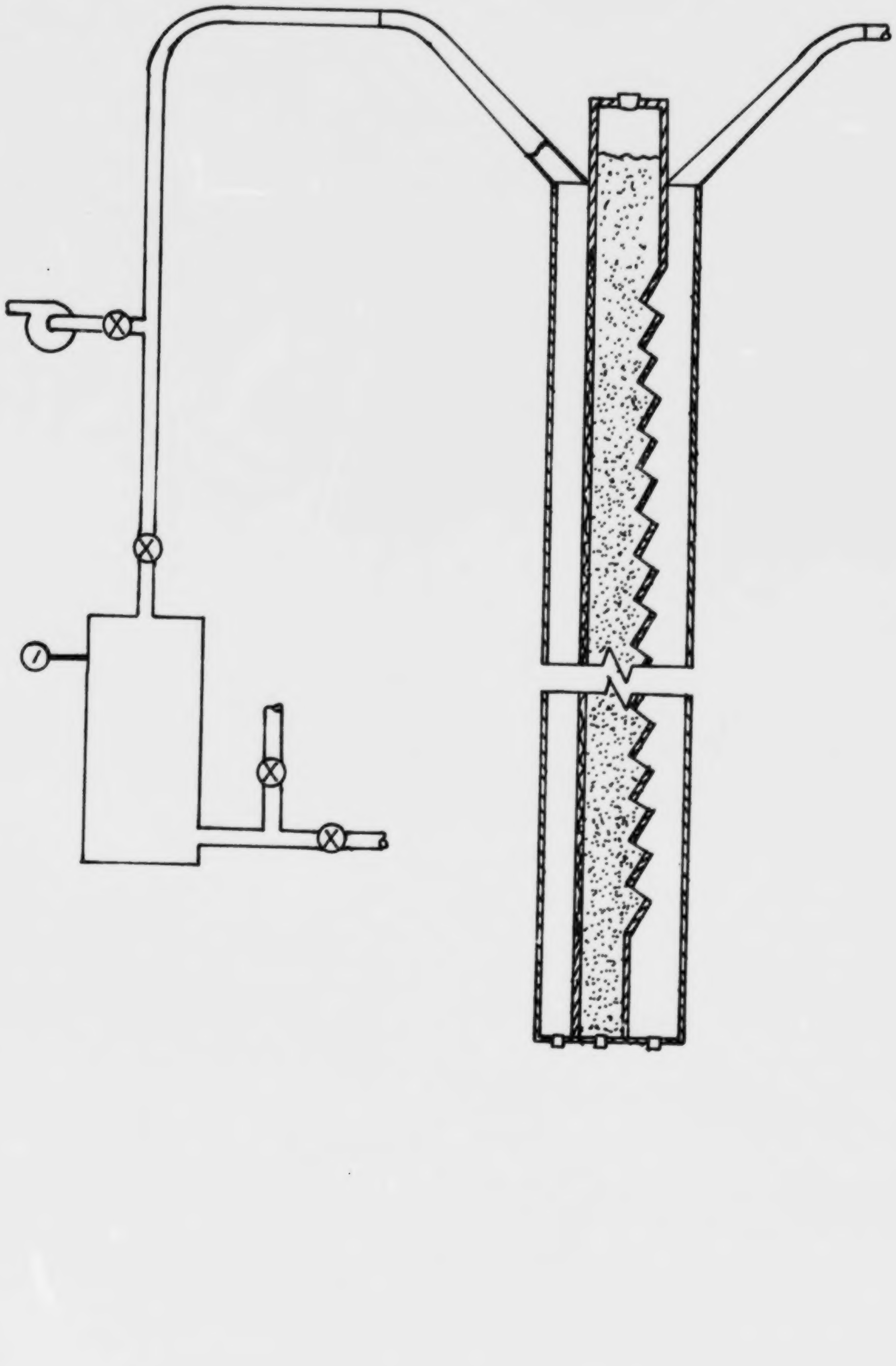
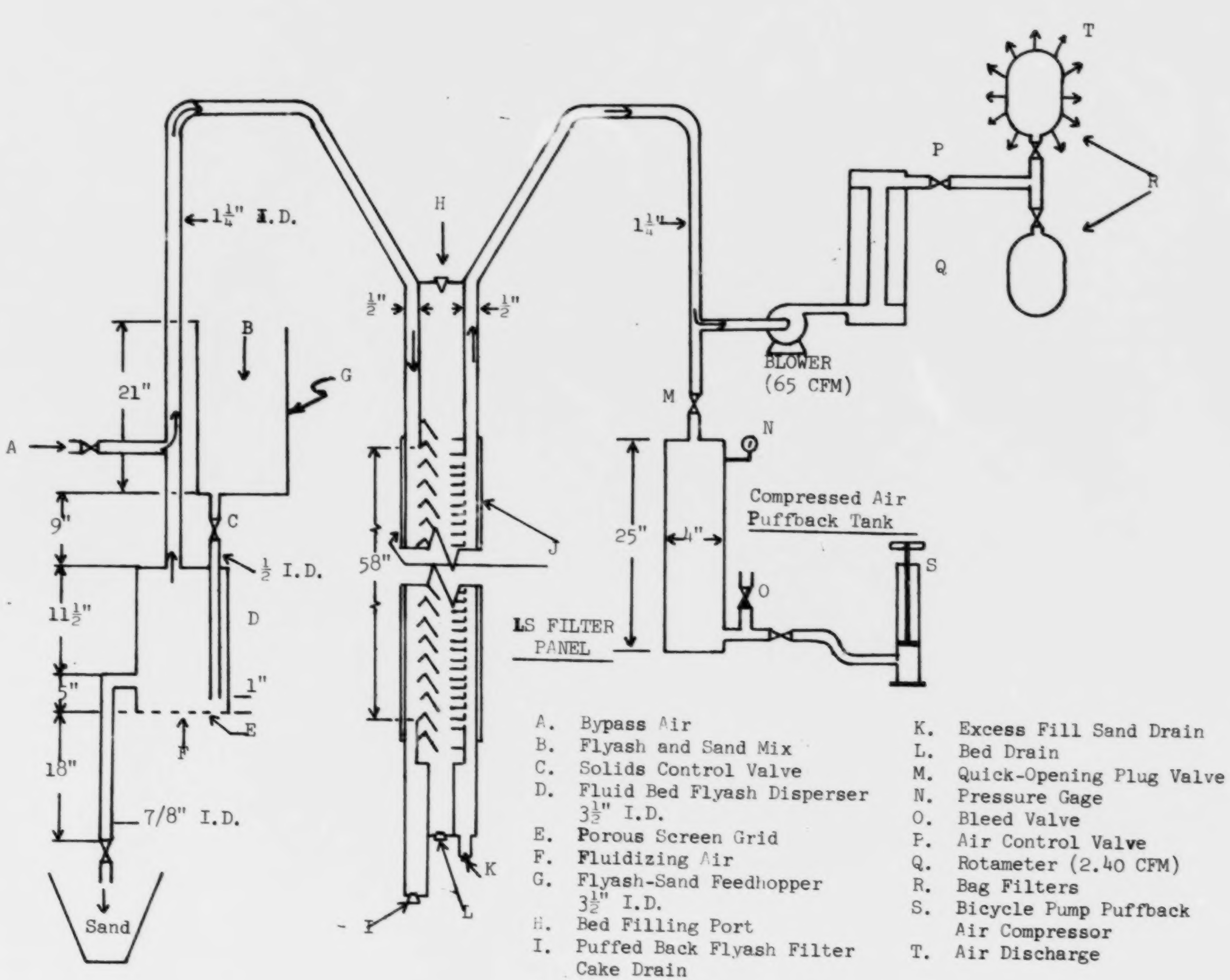

TIMER
Treated gas
Untreated Gas

Figure 9 Section of LS Panel (From Ref. 6)




T
P
R
Q
H
$1\frac{1}{4}$" I.D.
$1\frac{1}{4}$"
$\frac{1}{2}$"
$\frac{1}{2}$"
B
G
21"
A
BLOWER
(65 CFM)
M
N
9"
C
$\frac{1}{2}$ I.D.
J
58"
25"
4"
Compressed Air
Puffback Tank
S
O
$11\frac{1}{2}$"
D
LS FILTER
PANEL
5"
1"
E
F
18"
7/8" I.D.
K
Sand
I
L

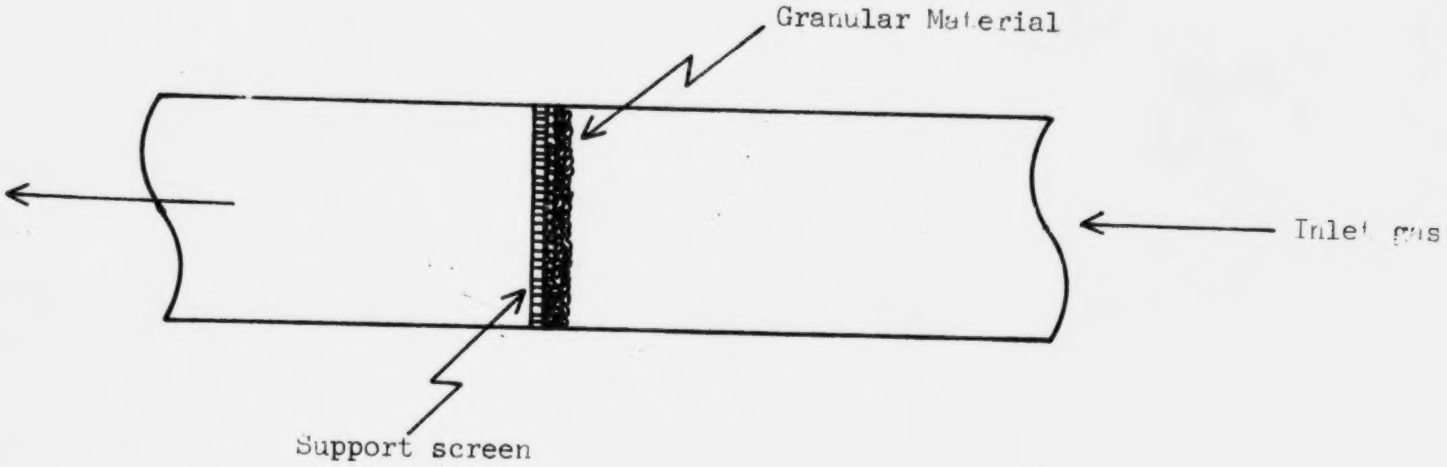
A. Bypass Air
B. Flyash and Sand Mix
C. Solids Control Valve
D. Fluid Bed Flyash Disperser $3\frac{1}{2}$" I.D.
E. Porous Screen Grid
F. Fluidizing Air
G. Flyash-Sand Feedhopper $3\frac{1}{2}$" I.D.
H. Bed Filling Port
I. Puffed Back Flyash Filter Cake Drain
J. Lucite Plates
K. Excess Fill Sand Drain
L. Bed Drain
M. Quick-Opening Plug Valve
N. Pressure Gage
O. Bleed Valve
P. Air Control Valve
Q. Rotameter (2.40 CFM)
R. Bag Filters
S. Bicycle Pump Puffback Air Compressor
T. Air Discharge

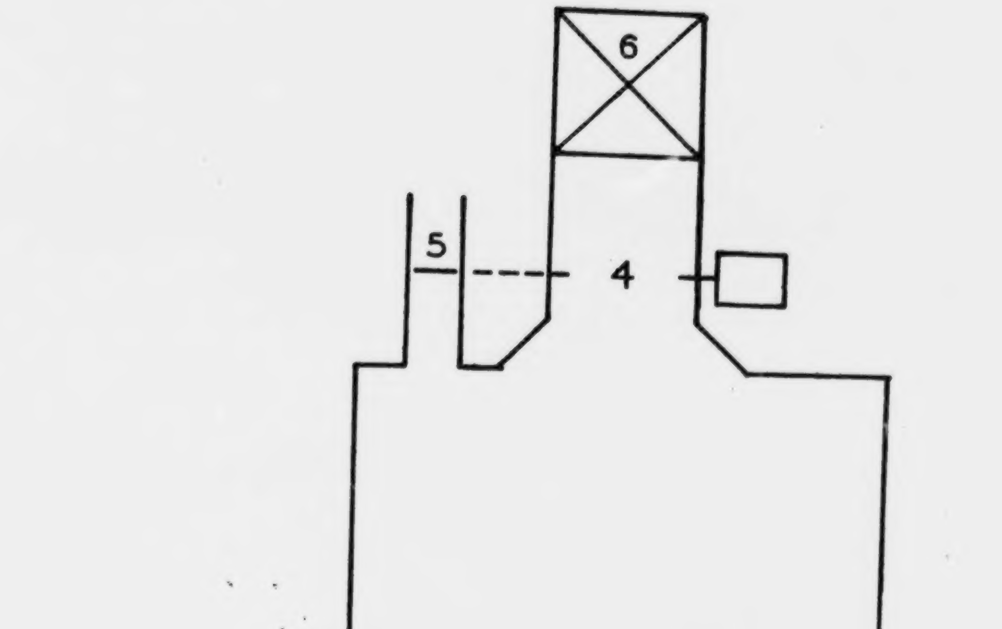
Figure 10 Experimental Arrangement for Tests on LS Filter

(From Material Supplied by A.M. Squires)

FIGURE 11

GRANGER FILTER CONCEPT


Granular Material
Inlet gas
Support screen



1. Inlet for dirty gas
2. Filter bed
3. Shaking motor
4. Shut-off flap
5. Scavenging air flap
6. Clean gas outlet
7. Hopper
8. Screw-type discharger

FIGURE 12

SECTION OF LURGI GRAVEL BED FILTER

FIGURE 13

COMPARISON BETWEEN INTERMITTENT FIXED BED AND STEADY MOVING BED

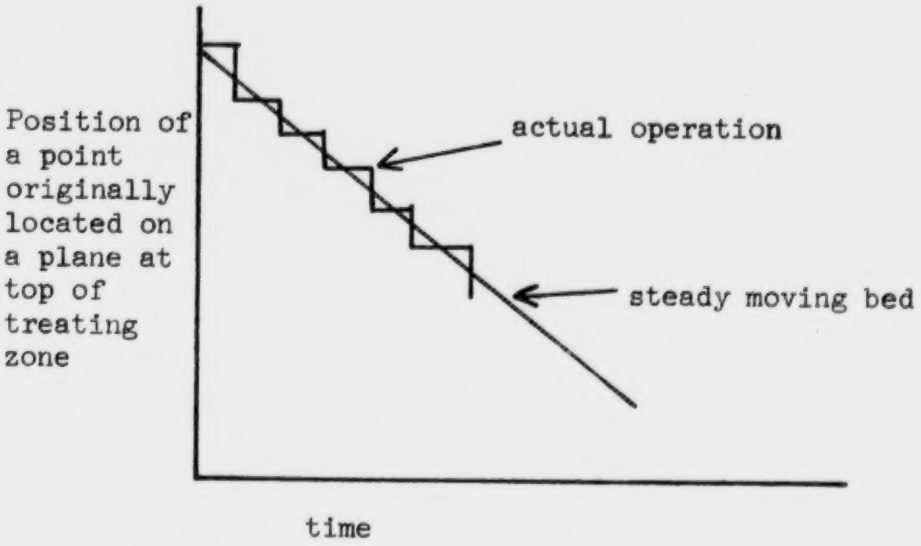

Position of a point originally located on a plane at top of treating zone
actual operation
steady moving bed
time

FIGURE 14

CROSSFLOW REACTOR VARIABLES

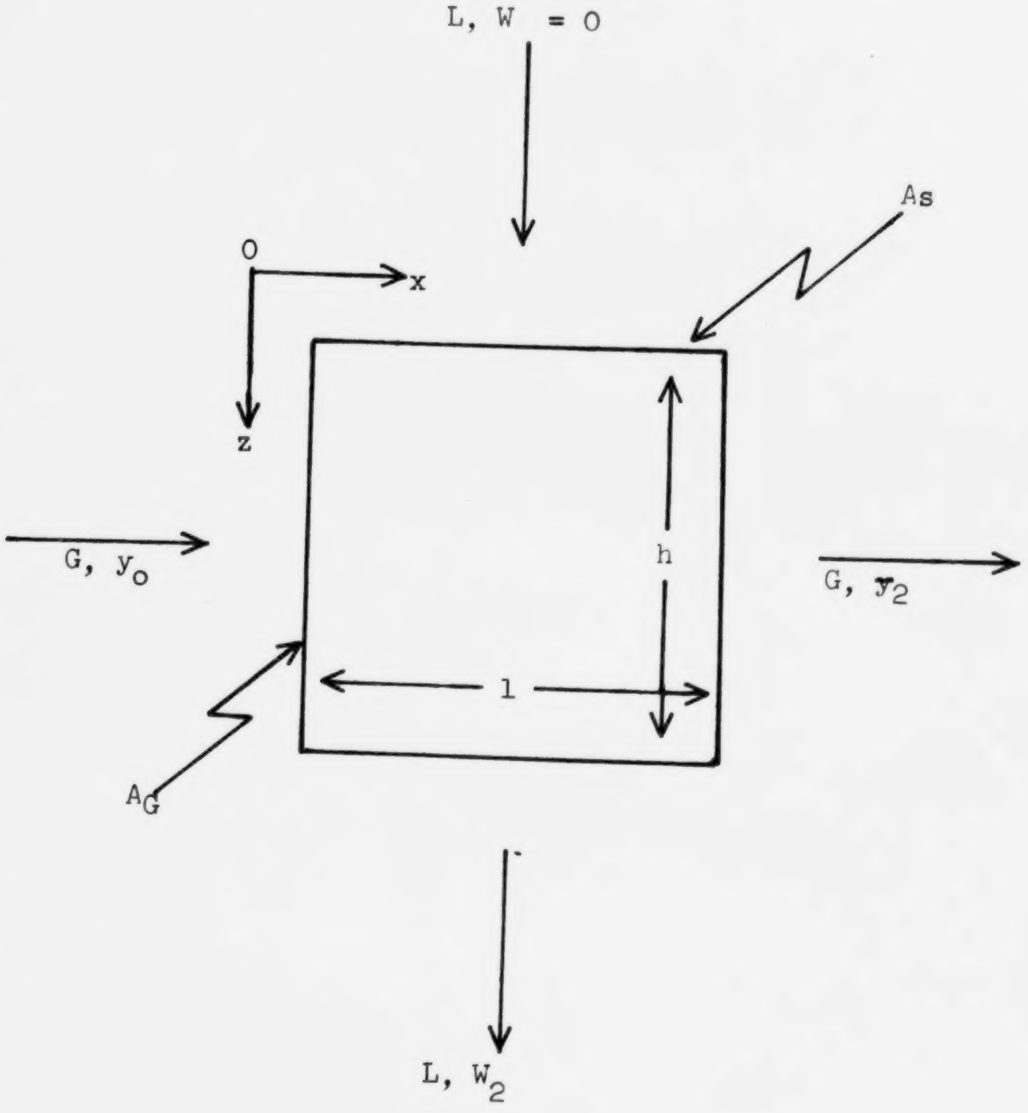
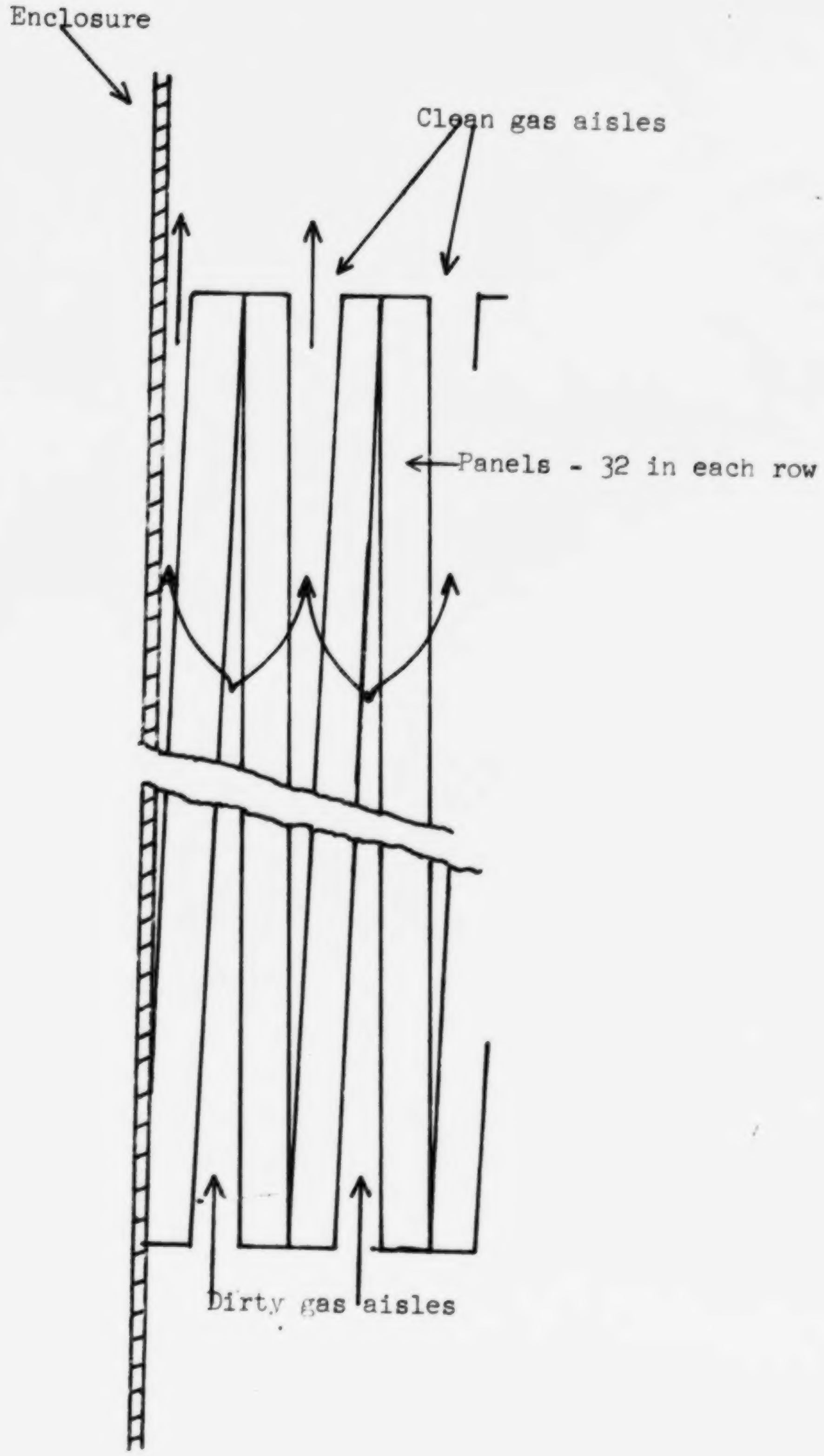
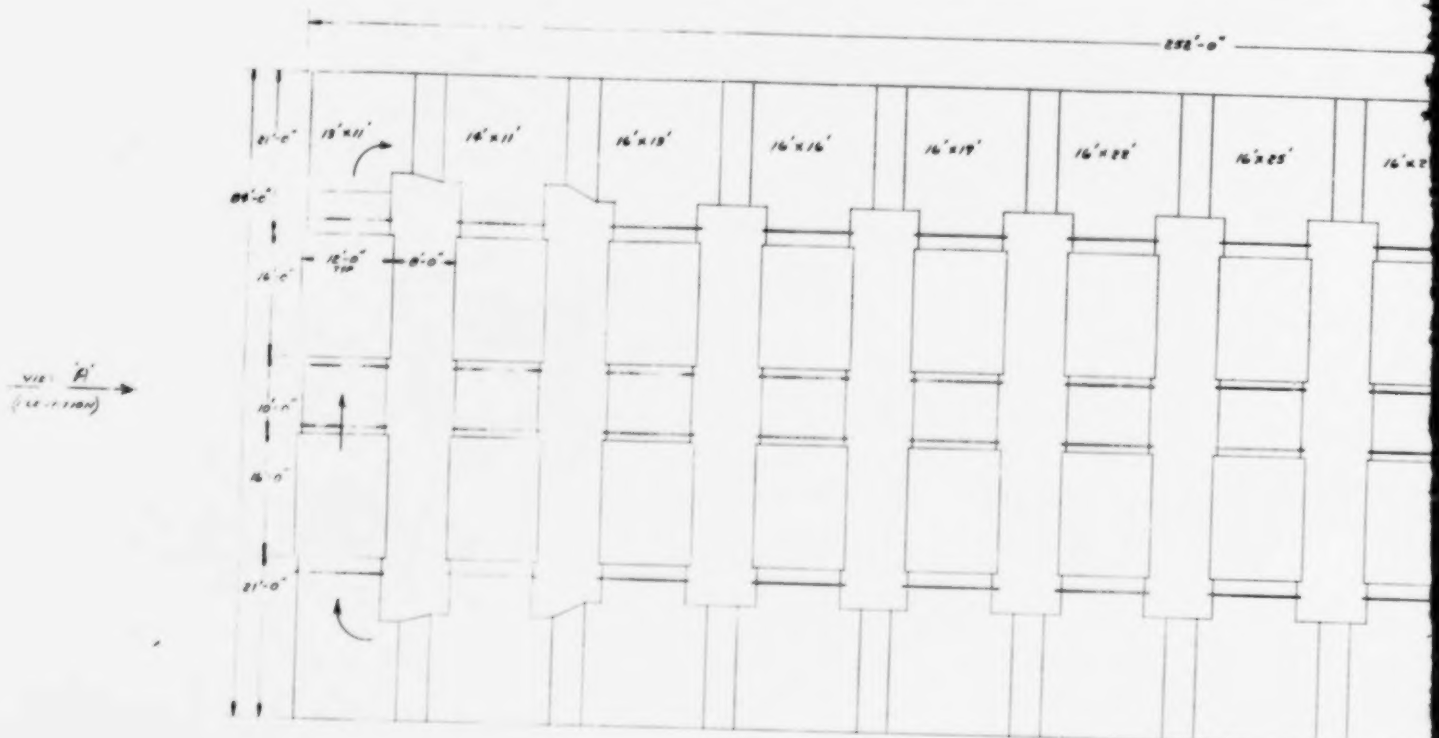

L, W = 0
As
0
x
z
G, y_o
h
G, y_2
l
A_G
L, W_2

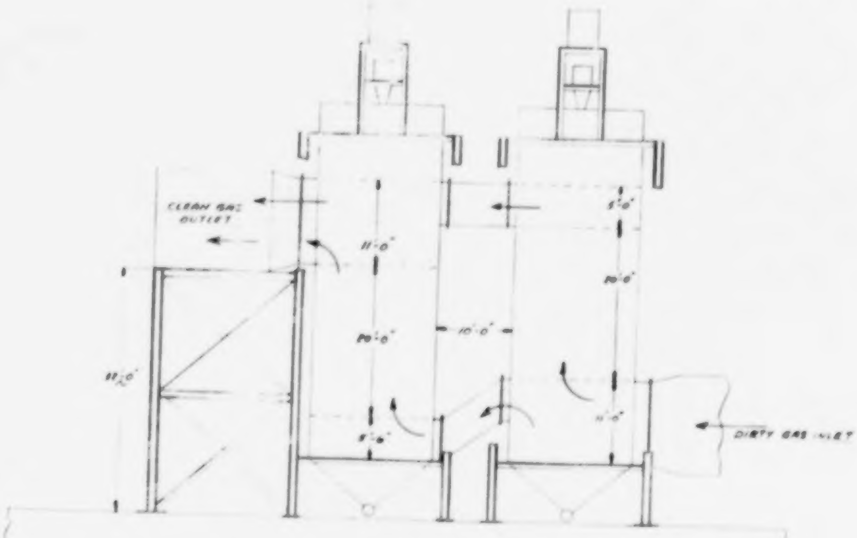
FIGURE 15

GAS FLOW PATTERNS IN ROW OF GSC PANELS


Enclosure
Clean gas aisles
Panels - 32 in each row
Dirty gas aisles


252'-0"
21'-0"
84'-0"
16'-0"
10'-0"
16'-0"
21'-0"
13'x11'
16'x11'
16'x13'
16'x16'
16'x19'
16'x22'
16'x25'
12'-0" TYP
8'-0"
VIEW 'A' (ELEVATION)

NOTE: ENTIRE COMPLEX COVERS AN AREA OF APPROX 88'x252' (24,16


CLEAN GAS OUTLET
11'-0"
20'-0"
5'-0"
20'-0"
10'-0"
11'-0"
8'-6"
27'-0"
DIRTY GAS INLET

VIEW 'A' (ELEVATION)

FIGURE 1

Layout and Elevation fo

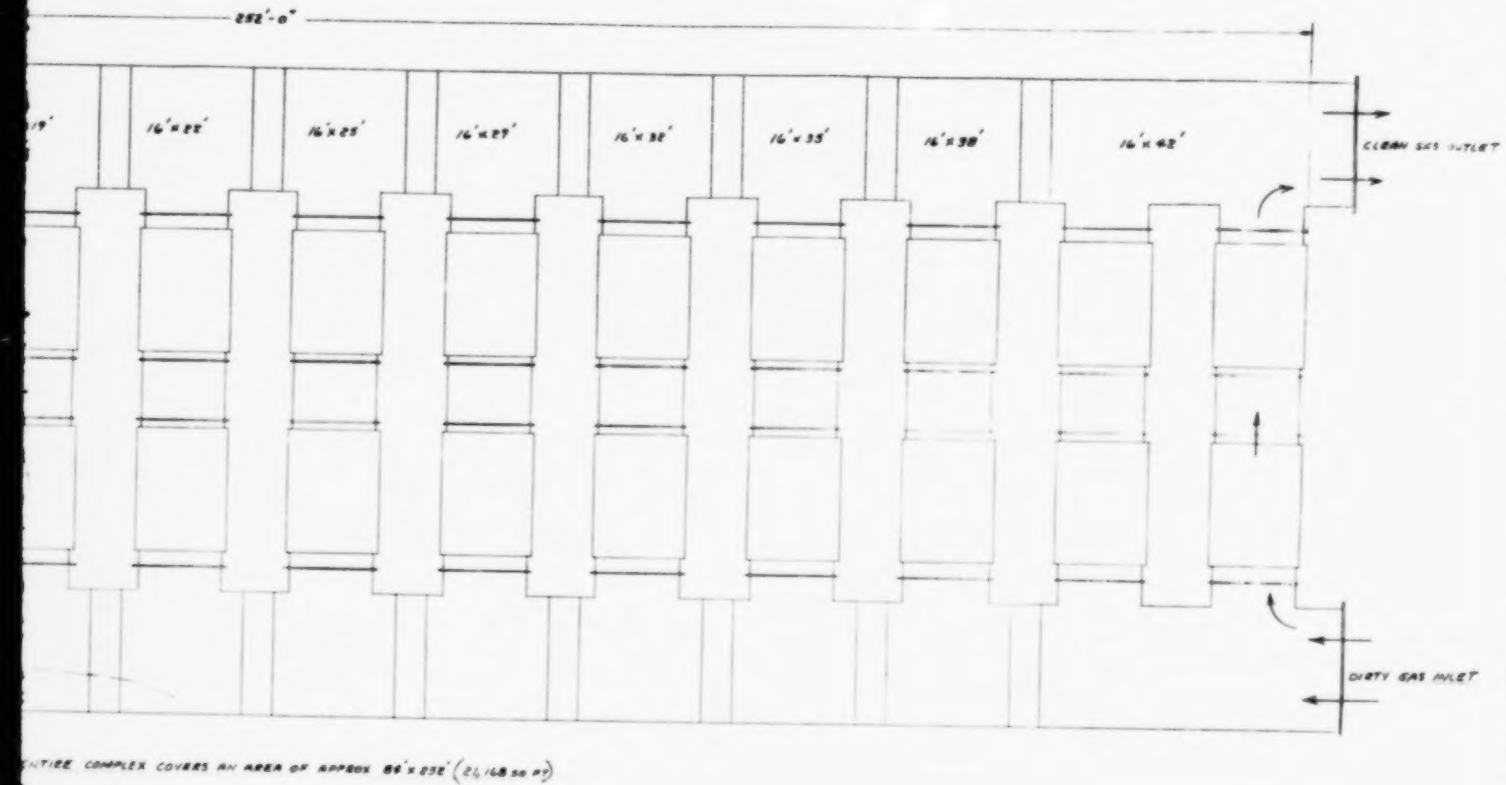

252'-0"
19'
16'×22'
16'×25'
16'×27'
16'×32'
16'×35'
16'×38'
16'×42'
CLEAN GAS OUTLET
DIRTY GAS INLET
ENTIRE COMPLEX COVERS AN AREA OF APPROX 84'×252' (21,168 SQ FT)

FIGURE 17

Layout and Elevation for GSC Enclosures

LIST OF PARTS

FIGURE 17

TITLE
SCHEMATIC - PLAN VIEW -
G.S.C. MODULES
800 MW POWER STATION

E 22372 T440-378

Ⓑ

FIGURE 16

(Separate Attachment)

Schematic - Individual Module GSC Panel

FIGURE 18

(Separate Attachment)

Schematic - Individual Filter Module Crossflow

FIGURE 19

PLAN VIEW OF CROSSFLOW ENCLOSURES WITH REVISED DUCTING DIMENSIONS

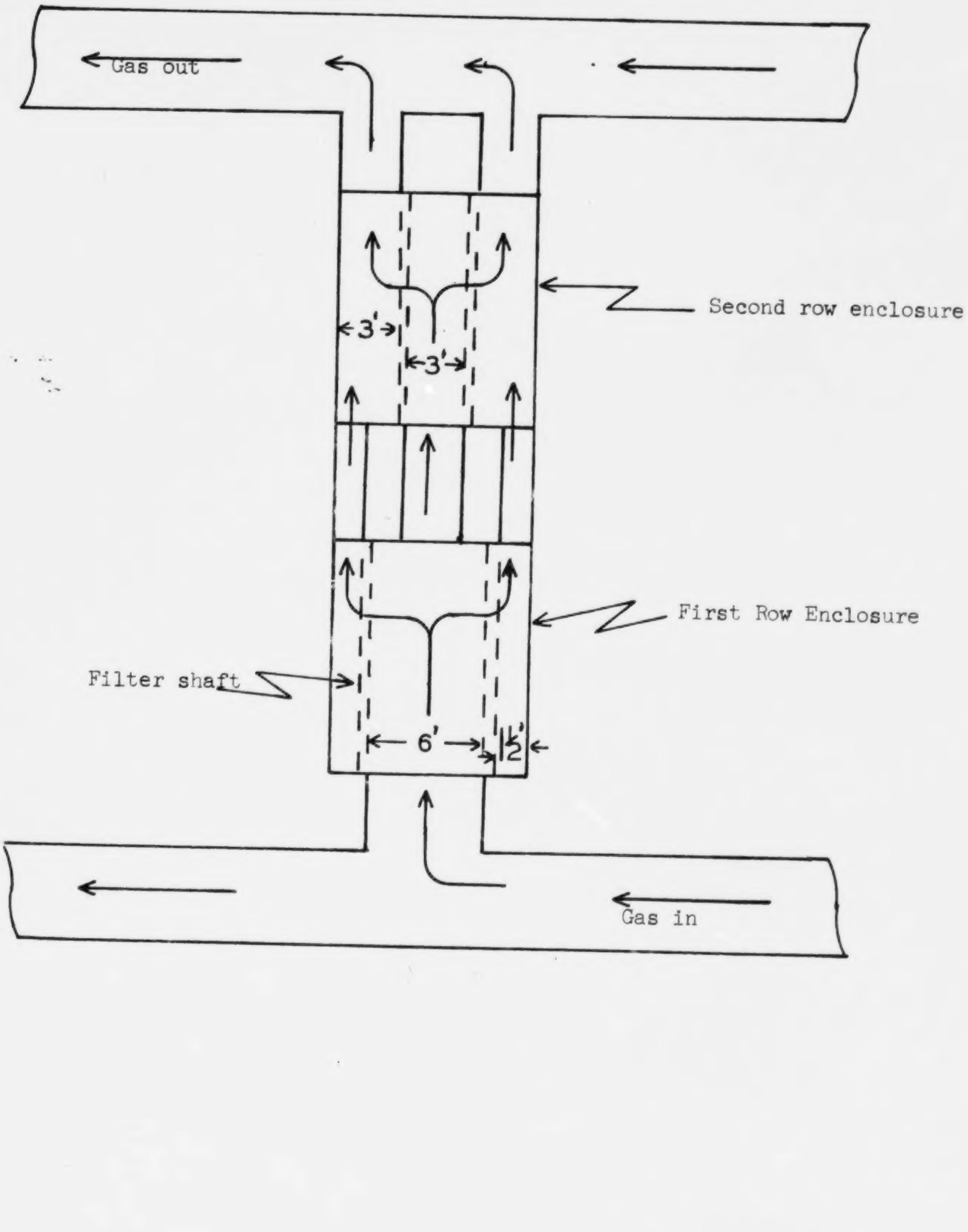
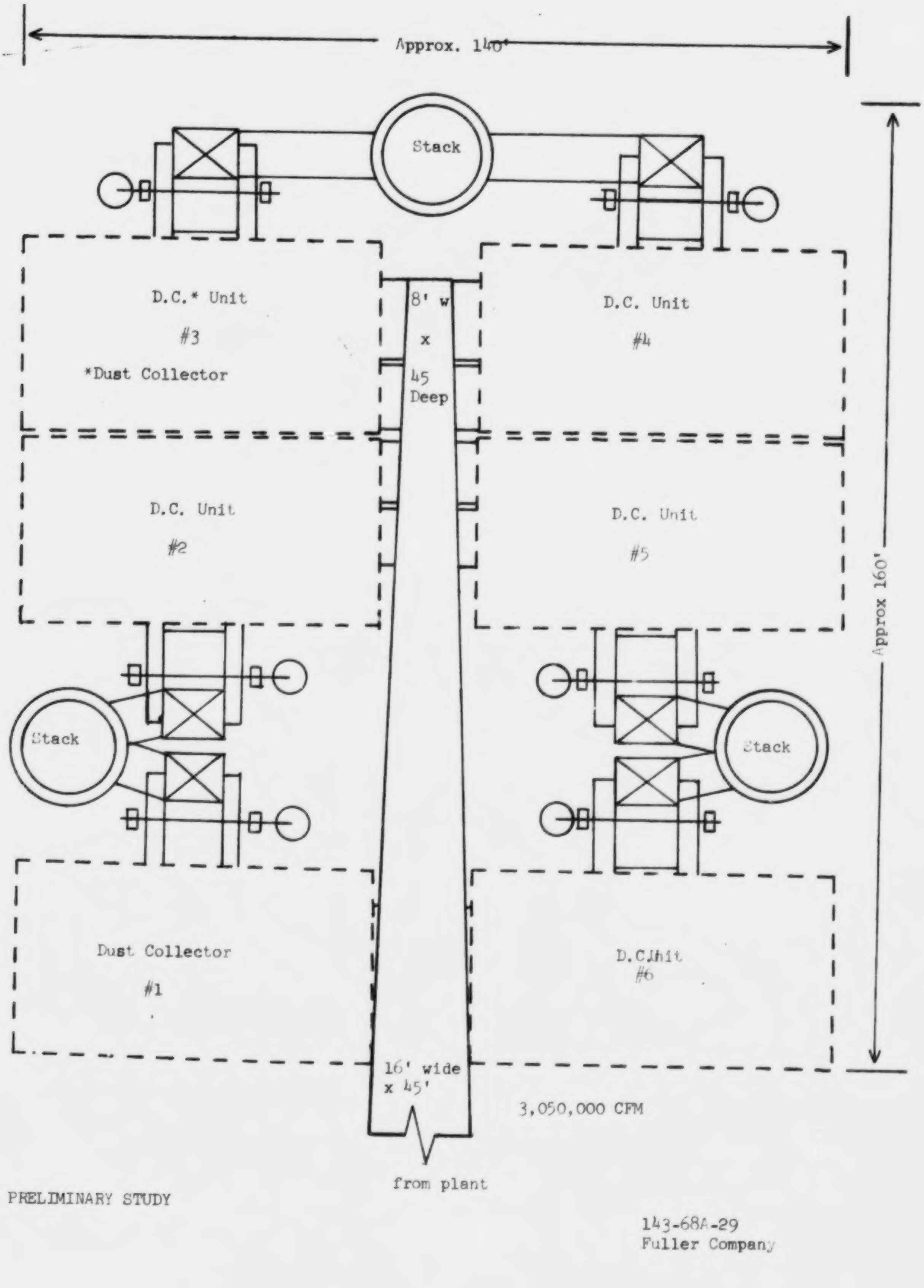

Gas out
Second row enclosure
3'
3'
First Row Enclosure
Filter shaft
6'
1/2'
Gas in

FIGURE 20

FULLER CO. LAYOUT


Approx. 140'
Stack
D.C.* Unit
#3
*Dust Collector
8' w
x
45
Deep
D.C. Unit
#4
D.C. Unit
#2
D.C. Unit
#5
Approx 160'
Stack
Stack
Dust Collector
#1
D.C.Unit
#6
16' wide
x 45'
3,050,000 CFM
from plant

PRELIMINARY STUDY

143-68A-29
Fuller Company

FIGURE 21

COMPARATIVE OPERATING COSTS

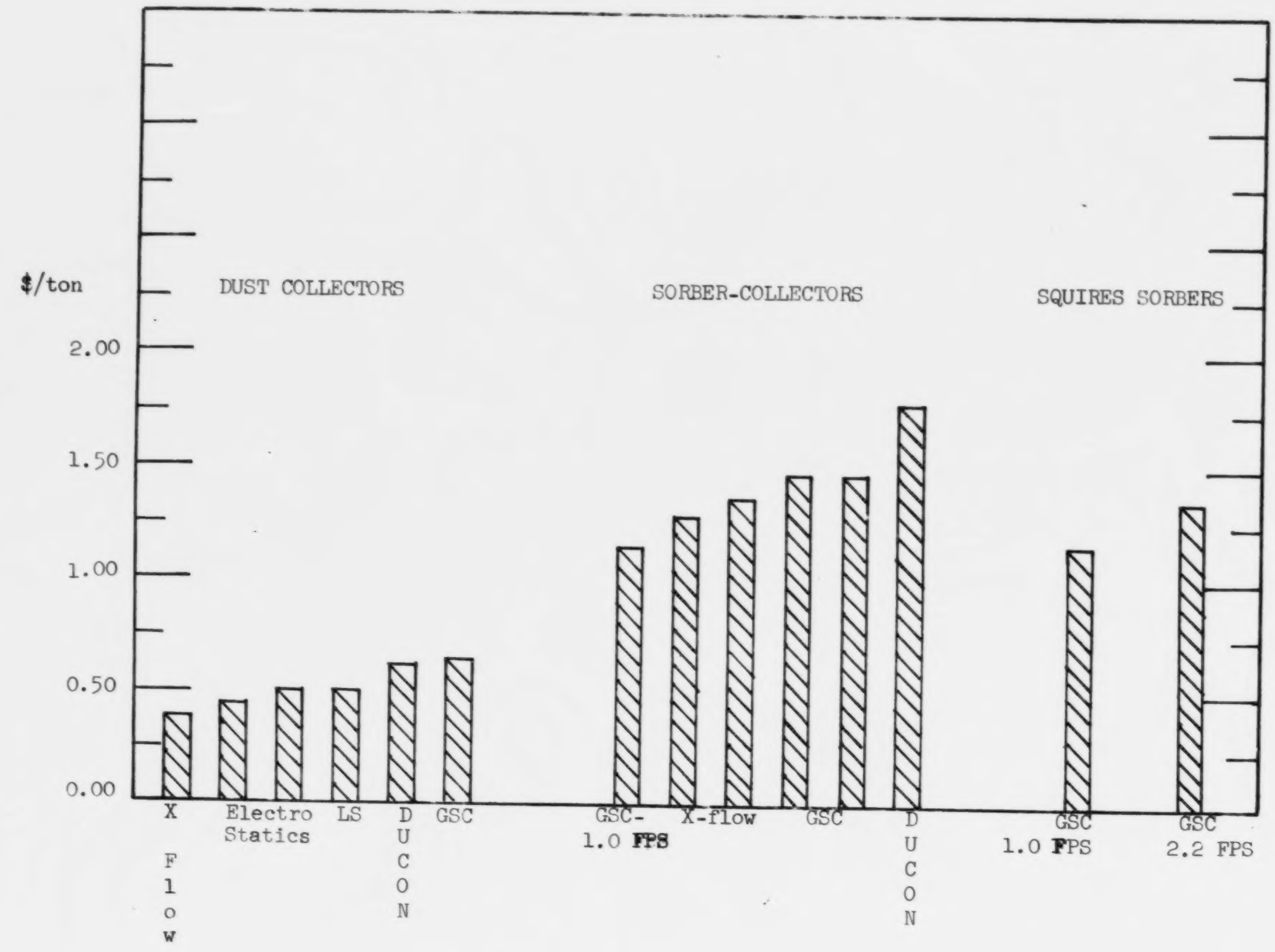
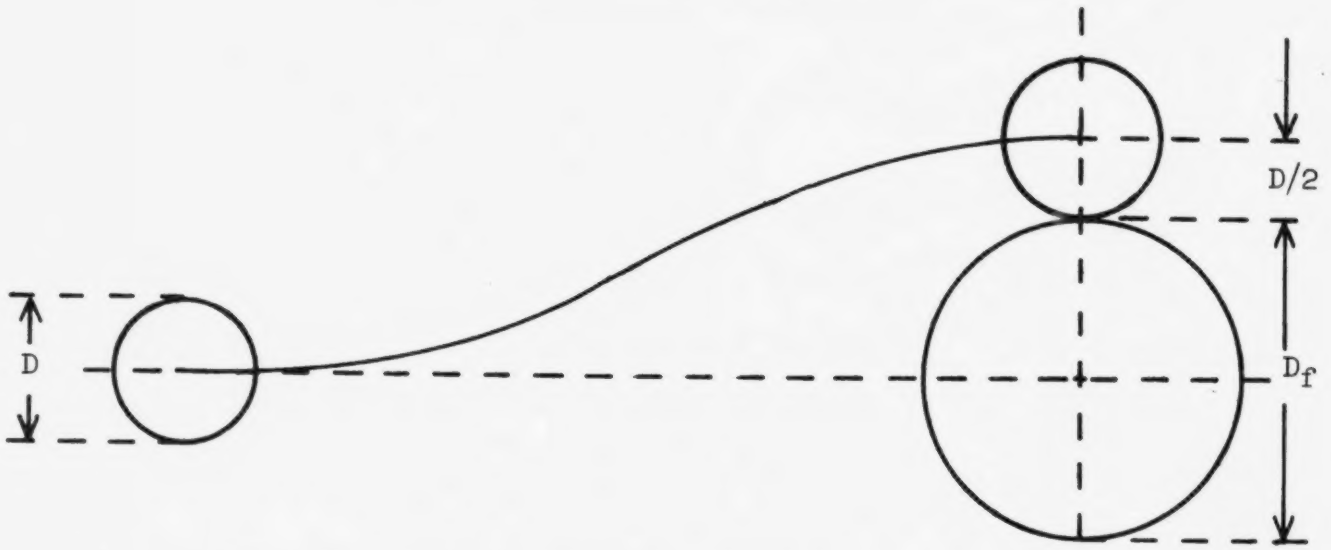


FIGURE 22

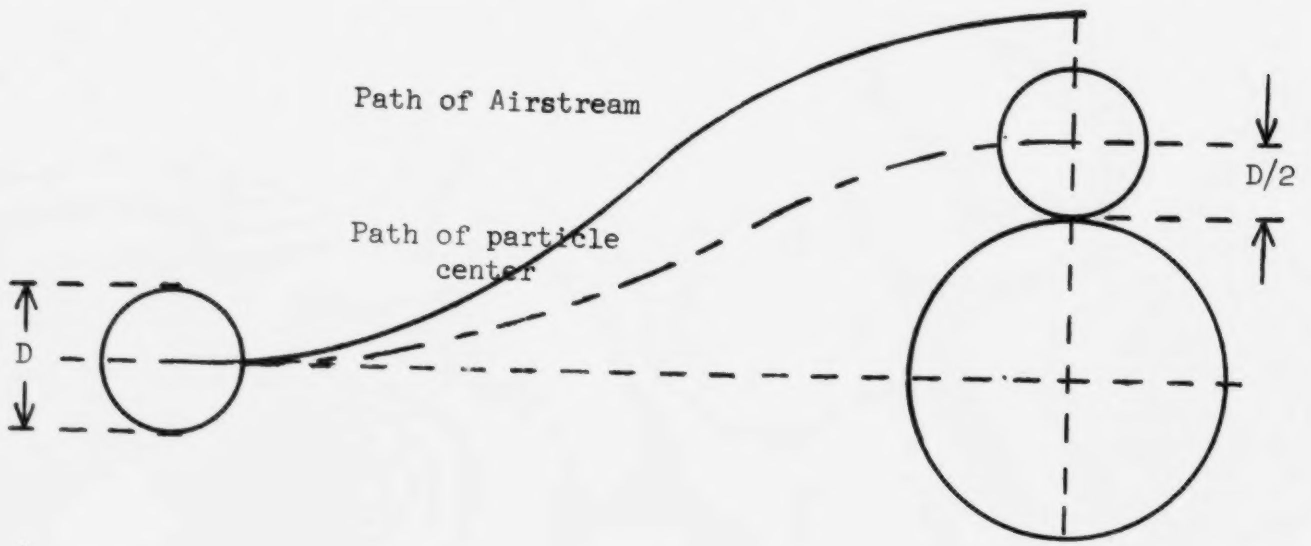
Collection by Direct Interception



DIRECT INTERCEPTION - PATH OF PARTICLE CENTER COINCIDENT WITH PATH OF AIRSTREAM

FIGURE 23

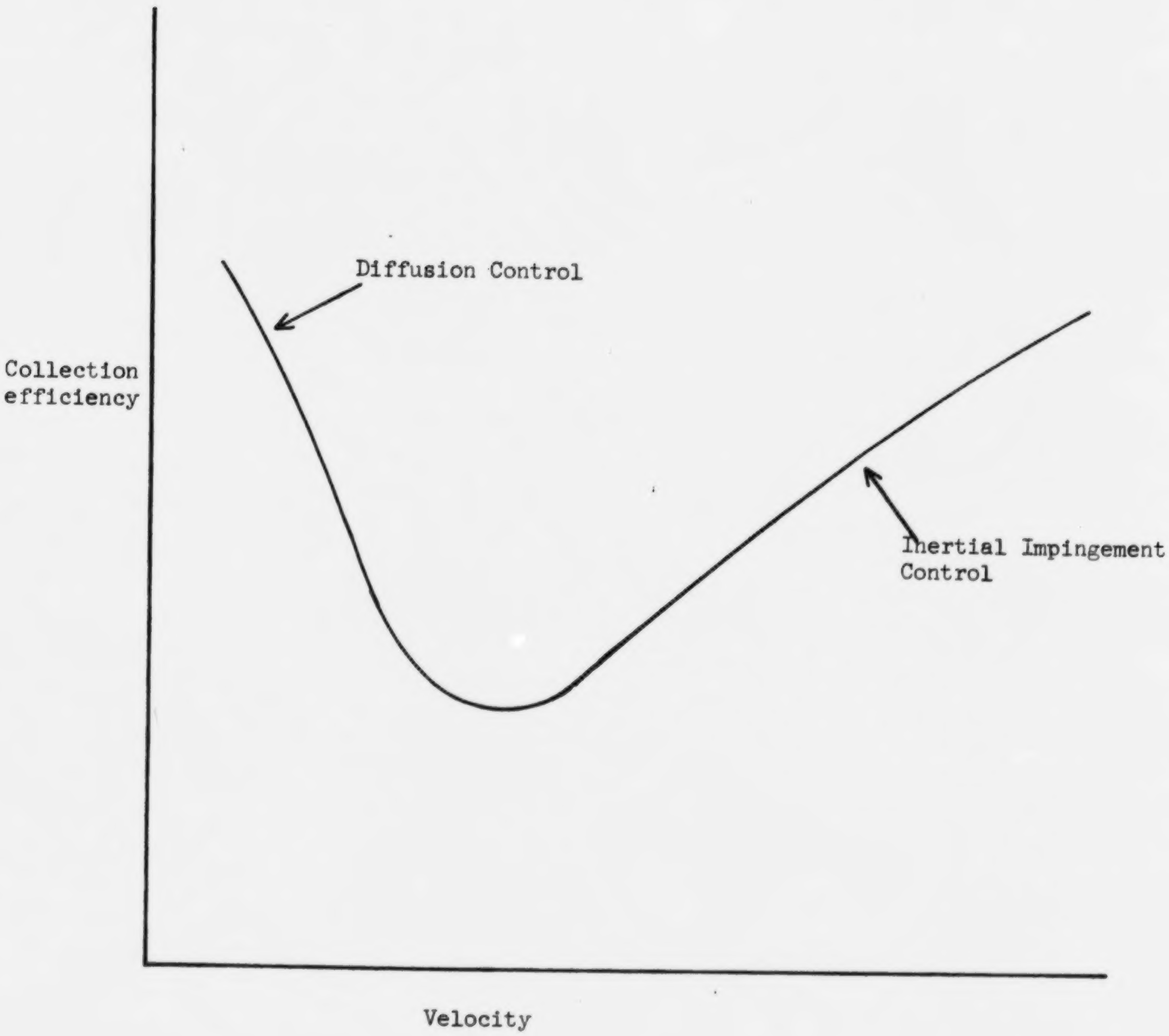
Collection by Impingement


Path of Airstream
Path of particle center
D
D/2

IMPINGEMENT - PATH OF PARTICLE CENTER NOT COINCIDENT WITH PATH OF AIRSTREAM

FIGURE 24

TREND OF COLLECTION EFFICIENCY WITH CHANGE IN CONTROLLING MECHANISMS



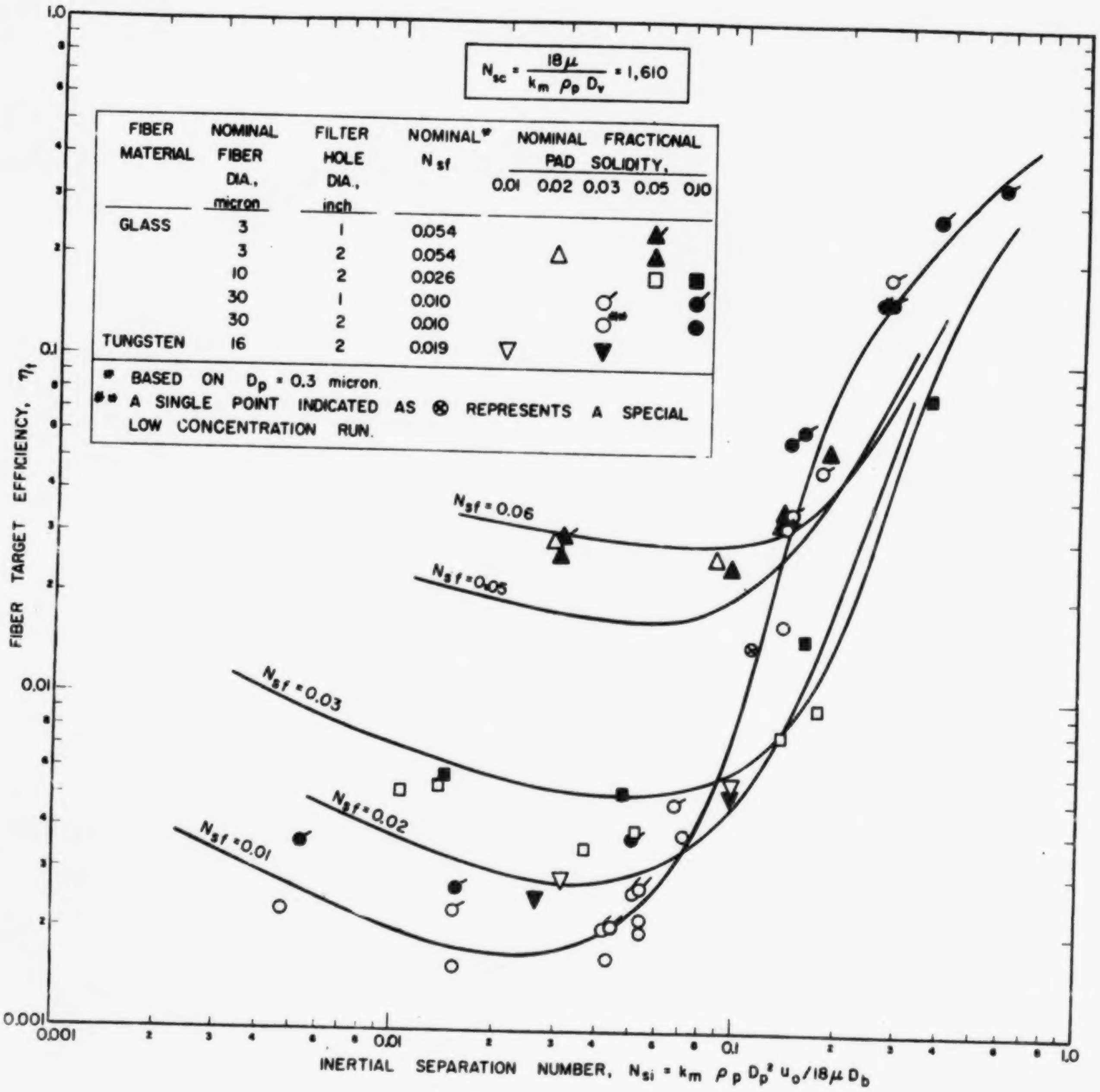


Figure 25 Target Efficiency as a Function of Inertial Separation Number for Aerosol L-1

FIGURE 26 COUNTERFLOW REACTOR VARIABLES

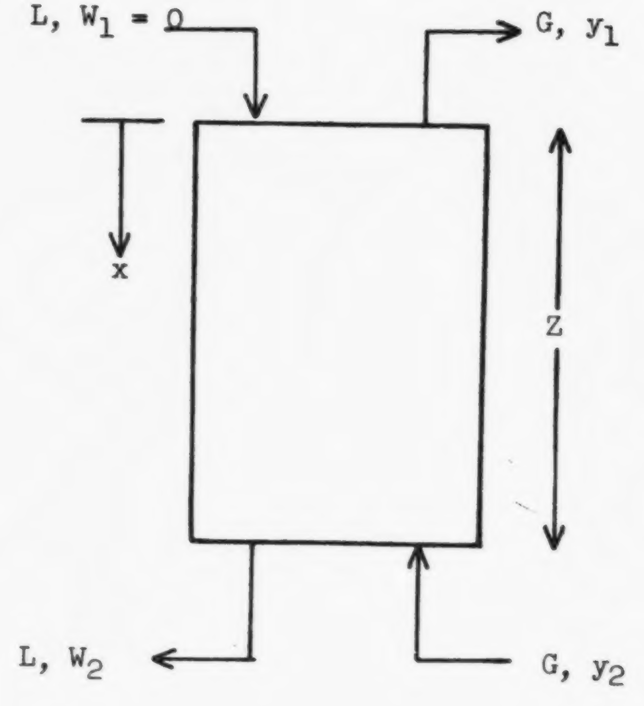

L, W_1 = 0
G, y_1
x
Z
L, W_2
G, y_2

FIGURE 27

CROSSFLOW REACTOR VARIABLES

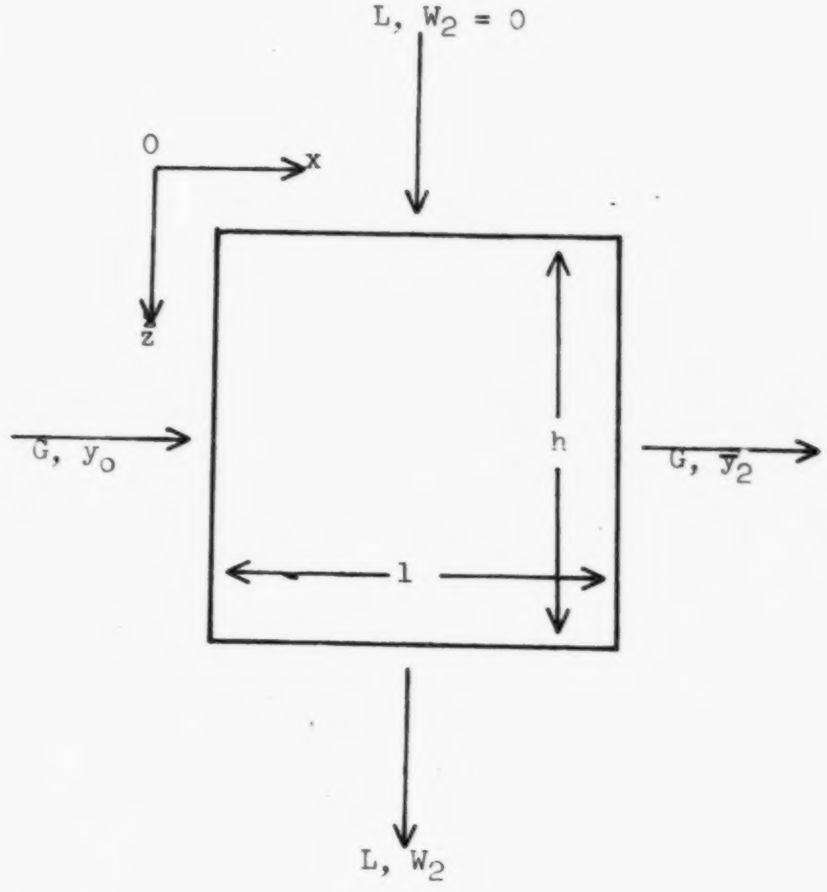

L, $W_2 = 0$
0
x
z
G, y_o
h
G, $\overline{y}_2$
l
L, W_2

FIGURE [illegible]

SORPTION PROFILE AT BOTTOM OF CROSSFLOW BED

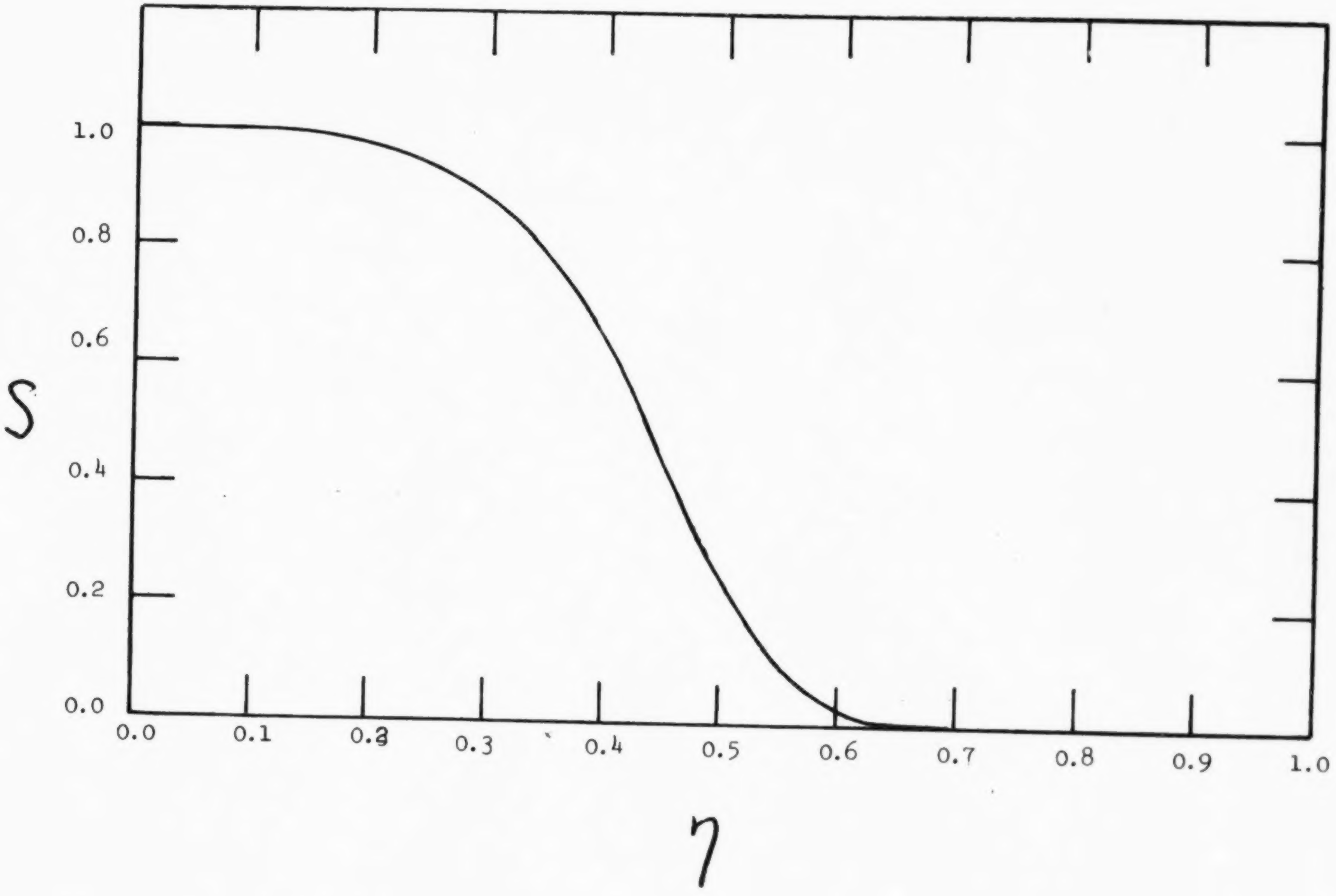
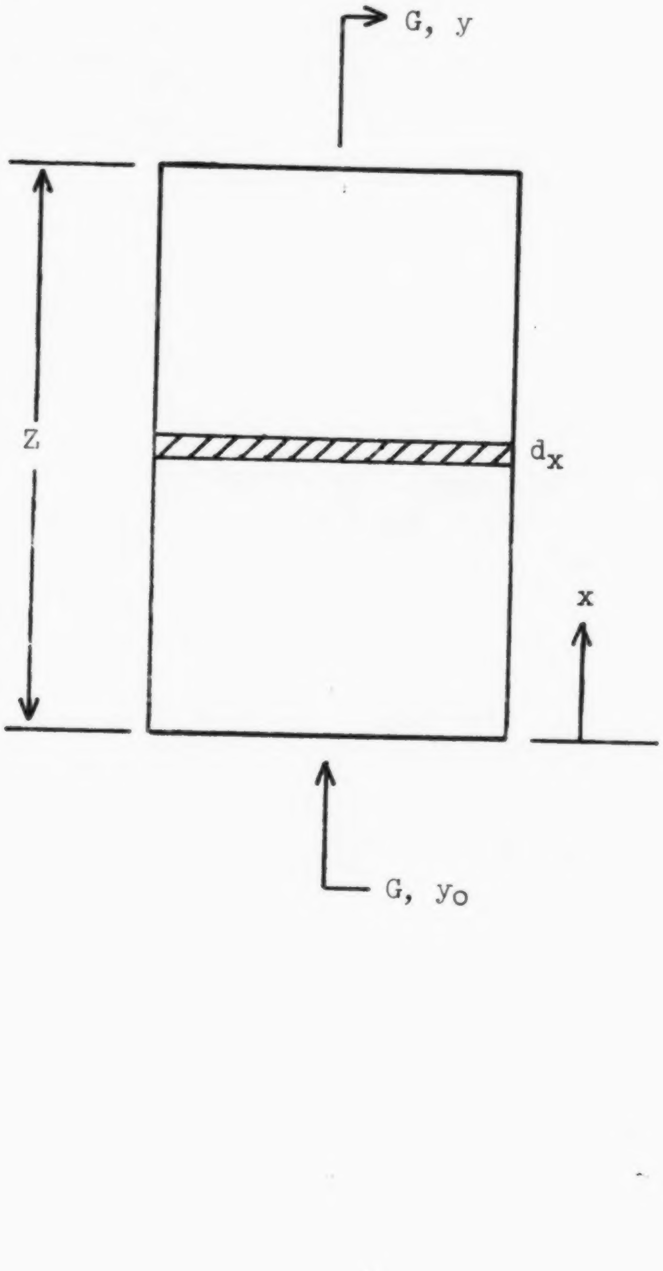
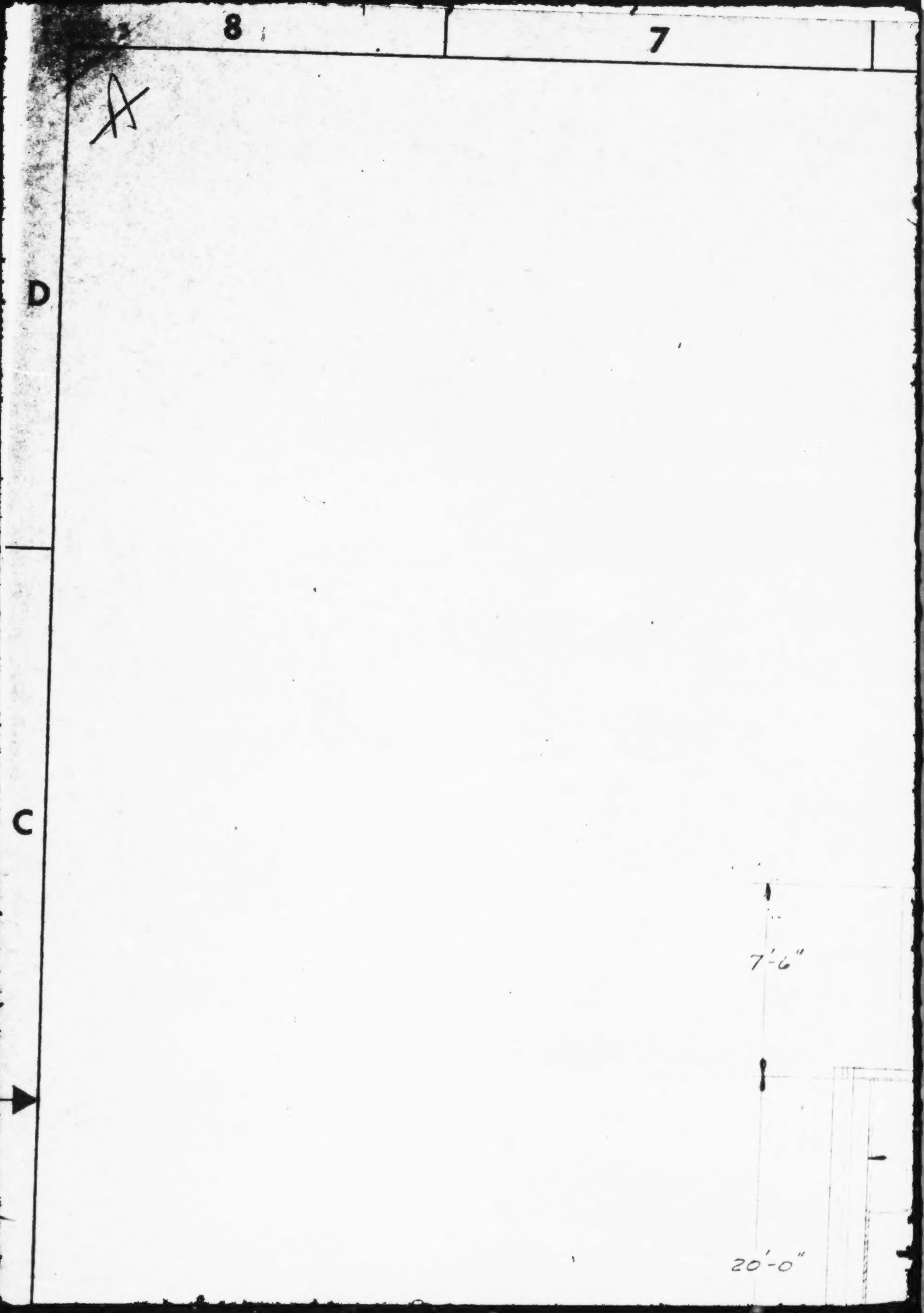


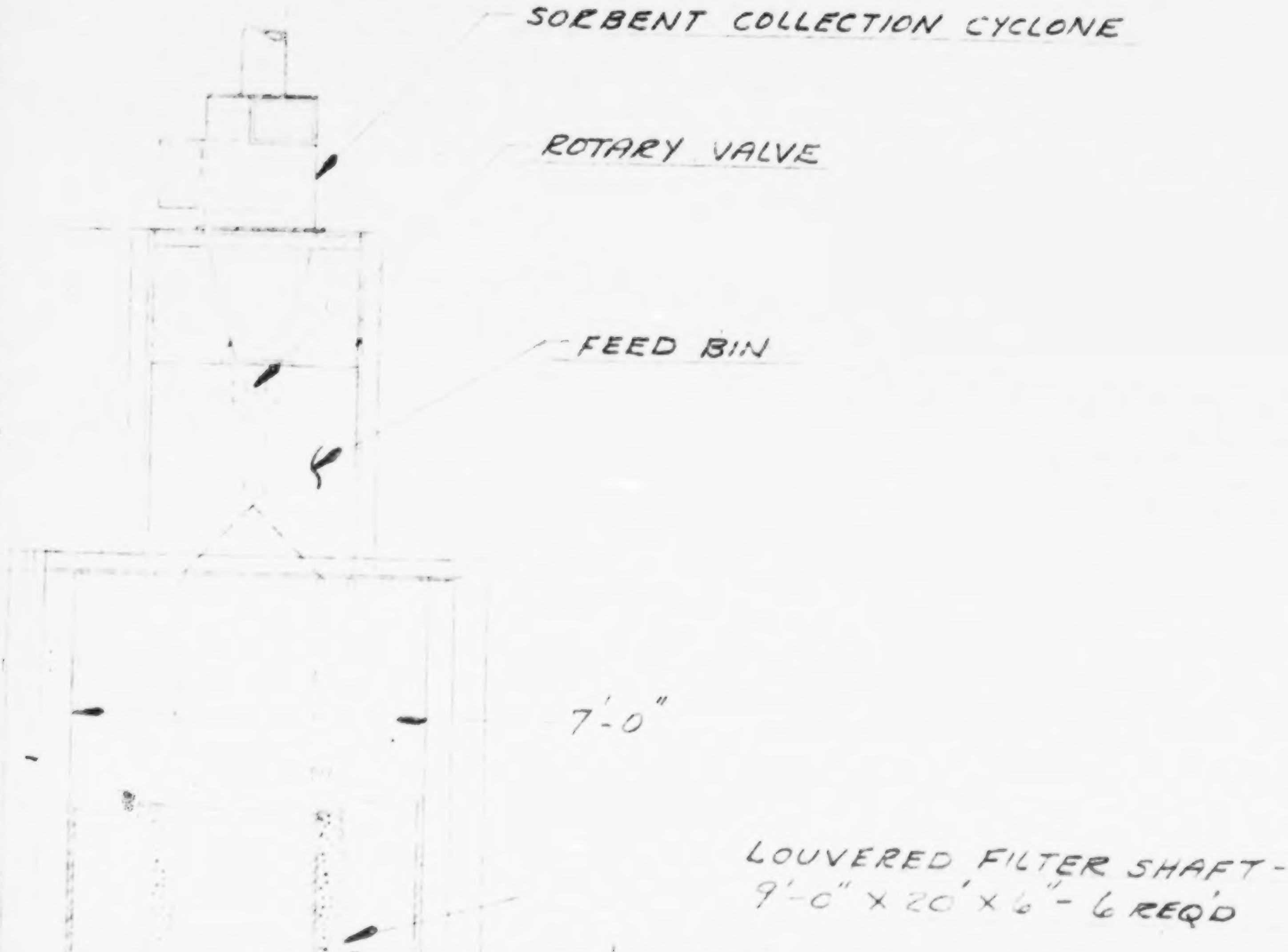
FIGURE 29

FIXED BED REACTOR



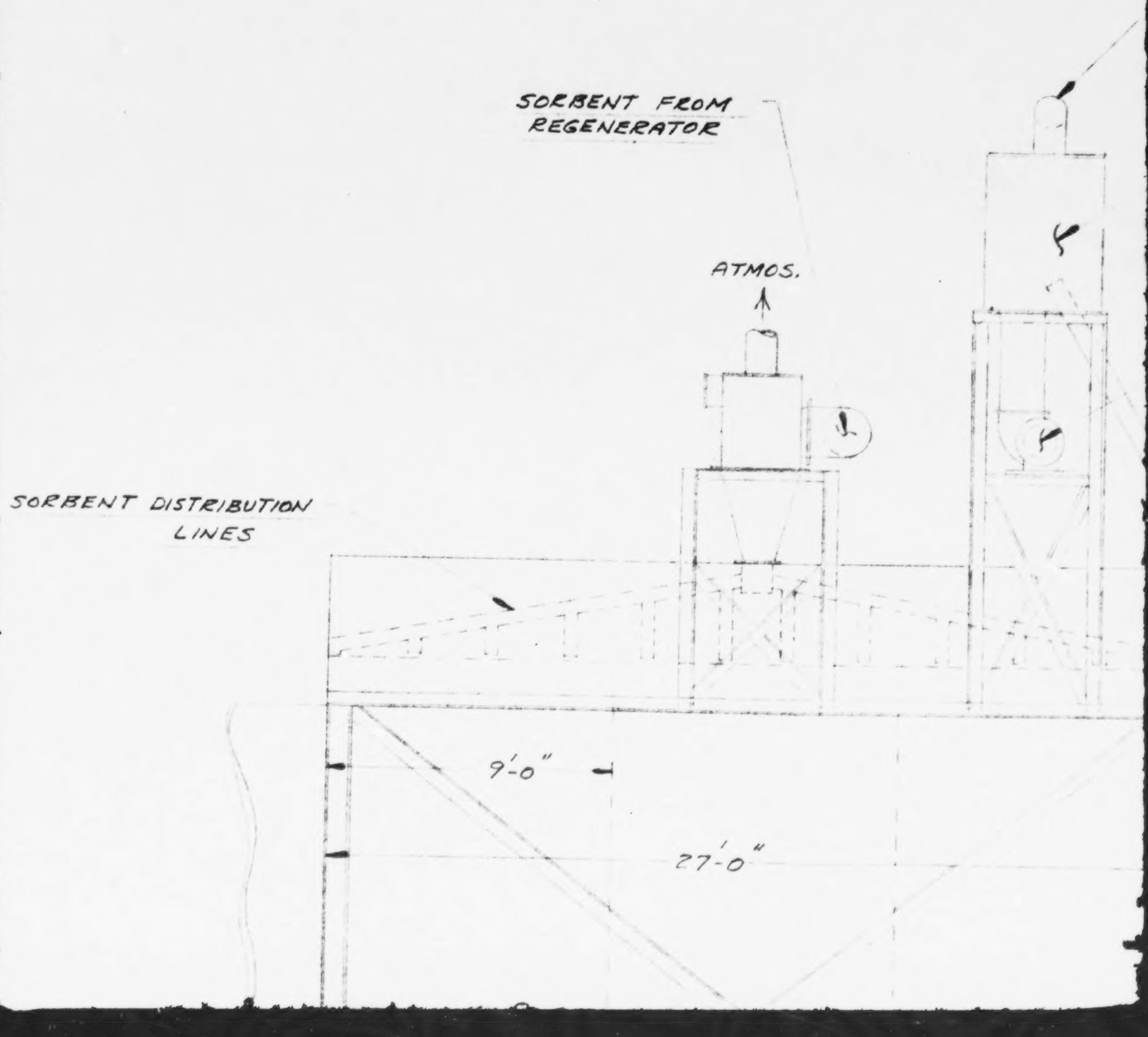

8
7
A
D
C
7'-6"
20'-0"

B


SORBENT COLLECTION CYCLONE
ROTARY VALVE
FEED BIN
7'-0"
LOUVERED FILTER SHAFT-
9'-0" x 20' x 6" - 6 REQ'D

SORBE

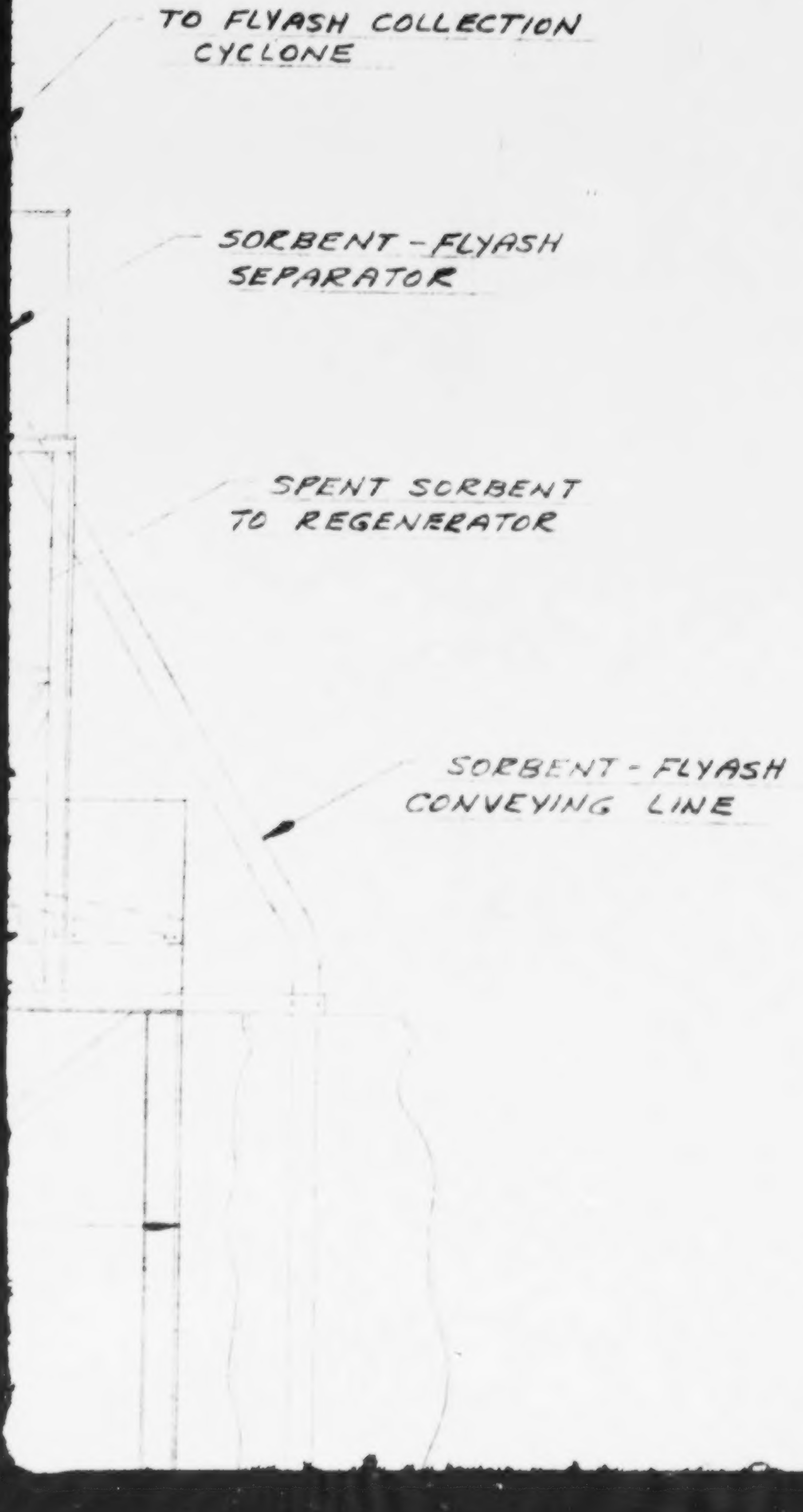
C

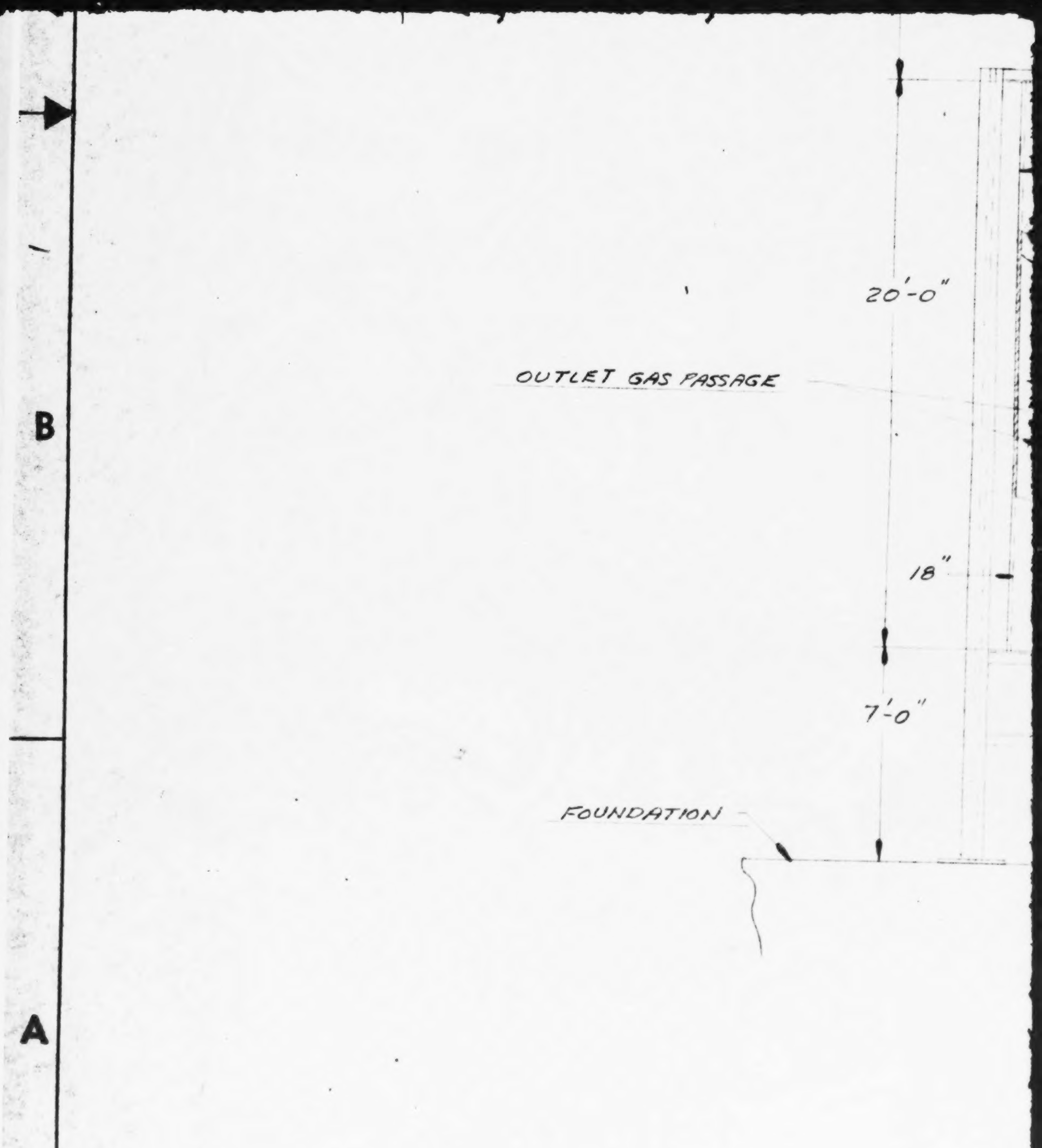

SORBENT FROM
REGENERATOR
ATMOS.
SORBENT DISTRIBUTION
LINES
9'-0"
27'-0"

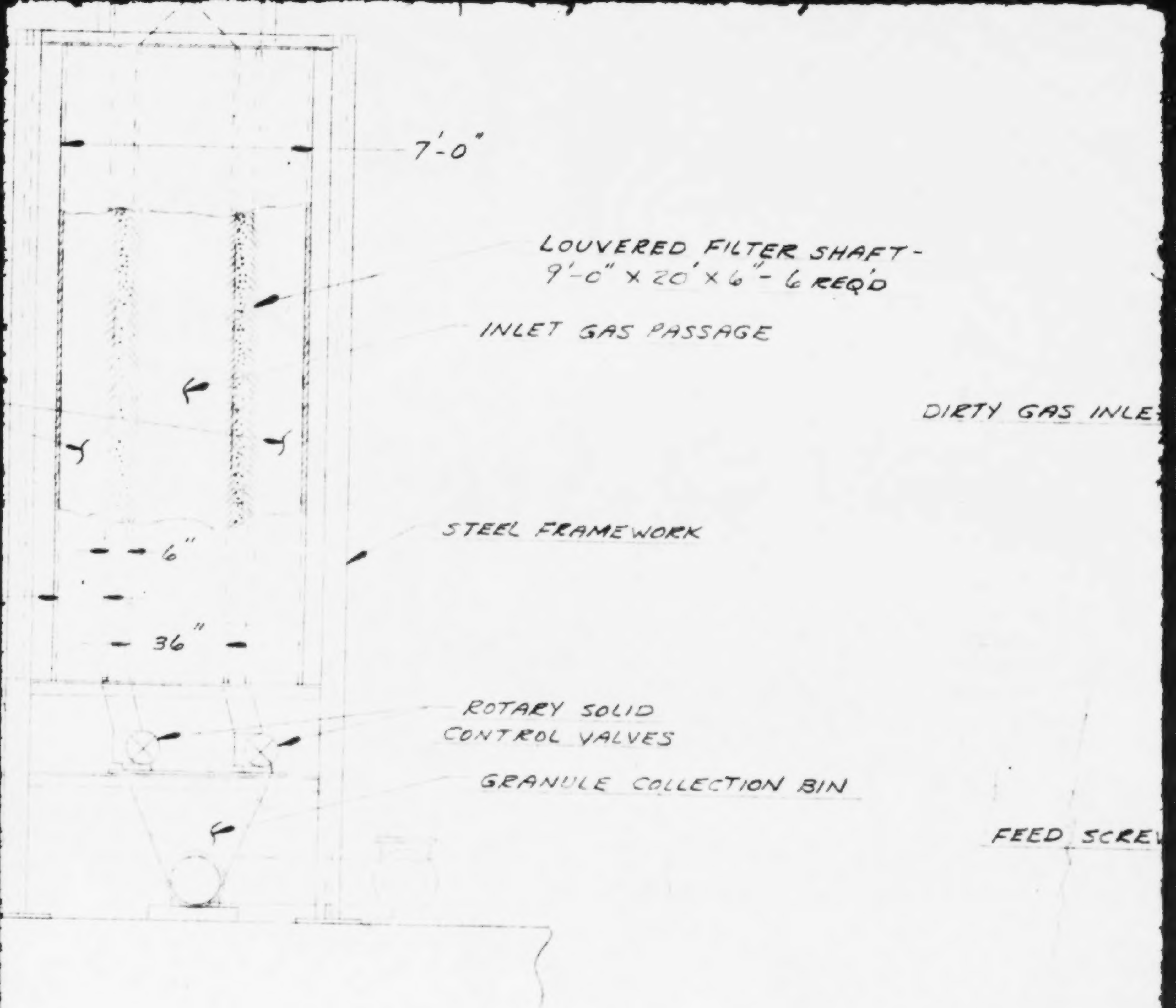
2 1

ENGR CHG	REVISIONS				
	ZONE	LTR	DESCRIPTION	DATE	APPROVED

D


TO FLYASH COLLECTION CYCLONE
SORBENT - FLYASH SEPARATOR
SPENT SORBENT TO REGENERATOR
SORBENT - FLYASH CONVEYING LINE
D
C


20'-0"
OUTLET GAS PASSAGE
18"
7'-0"
FOUNDATION
B
A
8
7

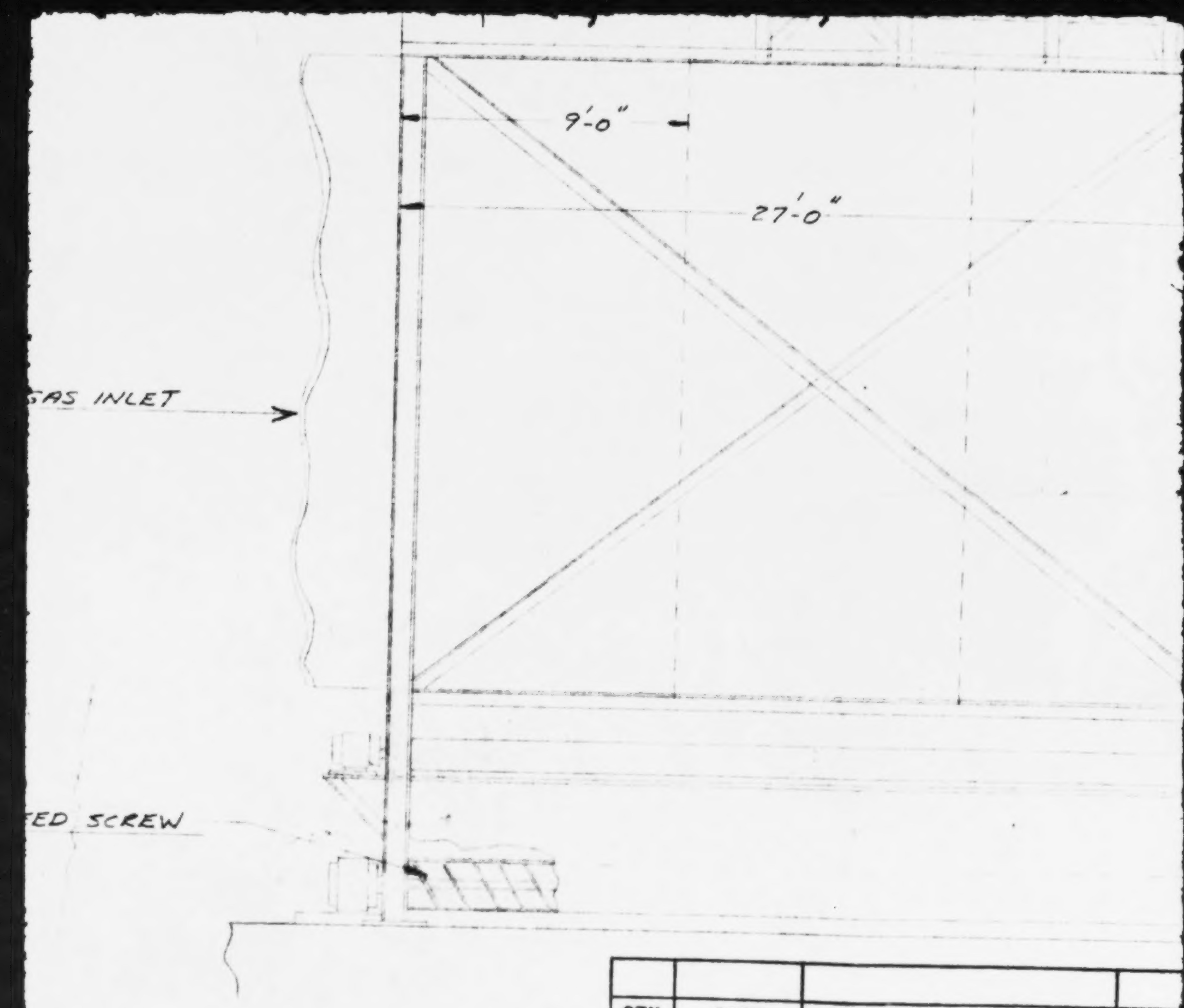

7'-0"
LOUVERED FILTER SHAFT -
9'-0" × 20' × 6" - 6 REQ'D
INLET GAS PASSAGE
DIRTY GAS INLE
STEEL FRAMEWORK
6"
36"
ROTARY SOLID
CONTROL VALVES
GRANULE COLLECTION BIN
FEED SCREW

PART CLASS:
DWG CLASS:
ASG BY:
APPD BY:

F

6

5



QTY REQD	CODE IDENT	PART OR IDENTIFYING NO.	
	←ASSY NO.		

DASH NO.	NEXT ASSY	USED ON	QTY PER NHA	END ITEM NUMBER	SERIAL NO.
	APPLICATION			EFFECTIVITY	

-UNLESS OTHERWISE SPECIFIED-

TOLERANCE AND DRAWING INTERPRETATION PER RAD-D143

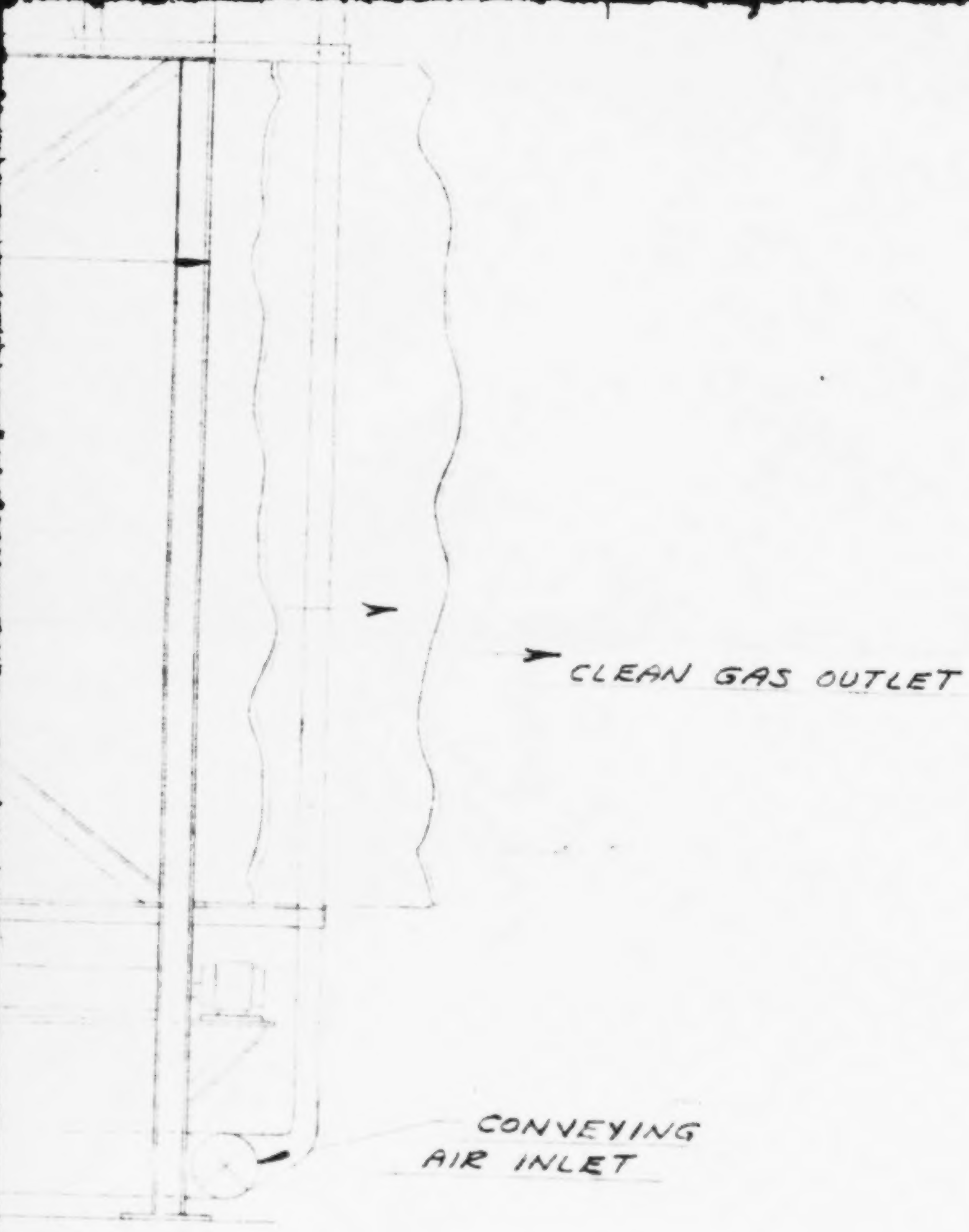
.X±.08 .XX±.04 .XXX±.020

MACHINED ANGLES±0° 30'

SHEET METAL BEND ANGLES±2°

RAD - P10001

SURFACE ROUGHNESS ✓



NOMENCLATURE OR DESCRIPTION	MATERIAL OR NOTE	SPECIFICATION	ZONE	ITEM NO.

LIST OF PARTS

CIFIED-
ING
-D143
X±.020
0'
ES±2°
✓

CONTR NO.		
RELEASE		
DWG APPD		
CHECKED		
DRAWN	Cote	9-6-68
DESIGN APPROVAL		
MANAGER		
SECT CH		
DESIGNER		

CORPORATION FIGURE 18
AVCO SPACE SYSTEMS DIVISION / LOWELL MASSACHUSETTS

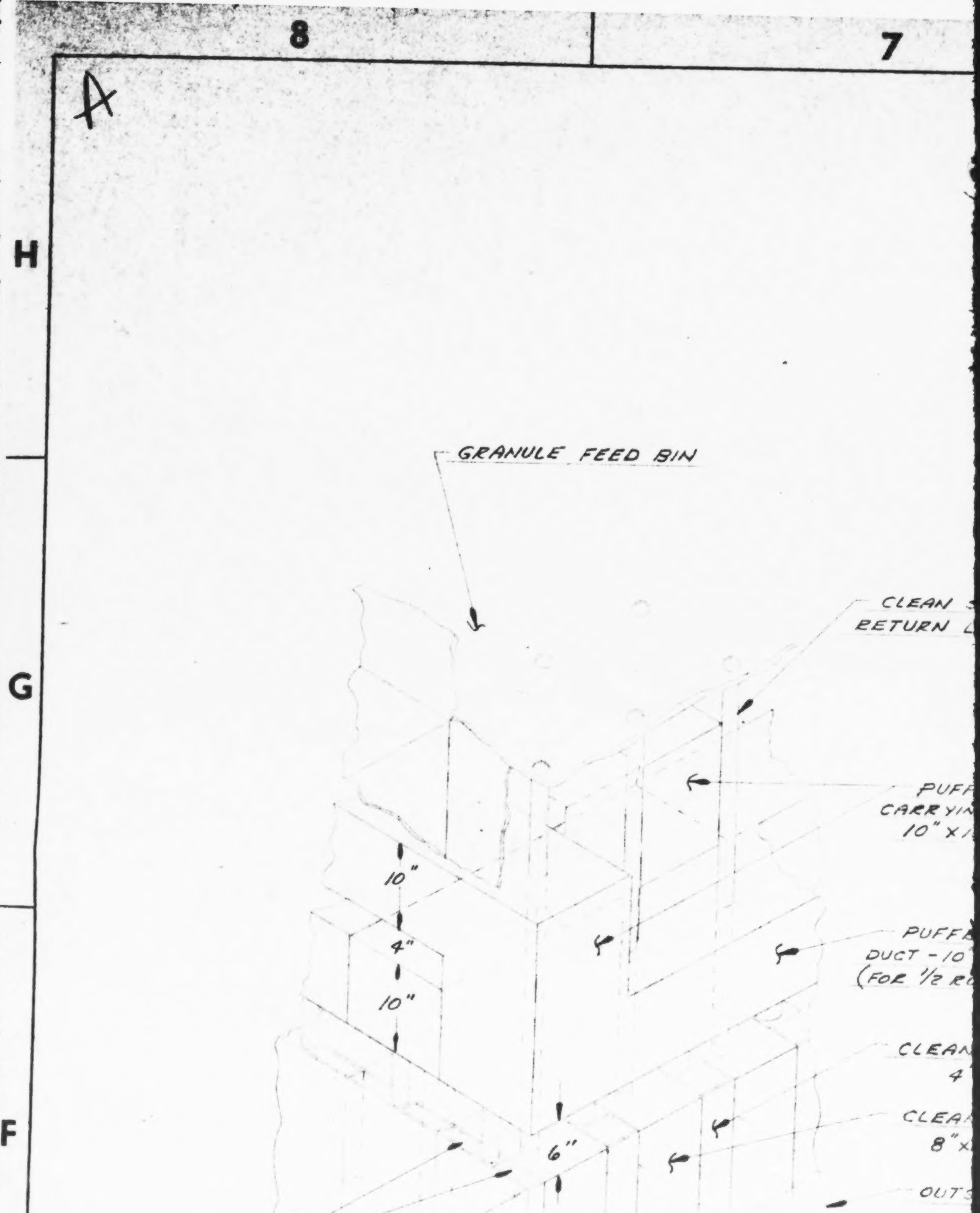
TITLE
SCHEMATIC - INDIVIDUAL FILTER MODULE - CAPACITY = 97,200 CFM

SIZE	CODE IDENT NO.	DWG NO.
D	22372	T440-376

SCALE 1/4"=1'-0" WT SHEET

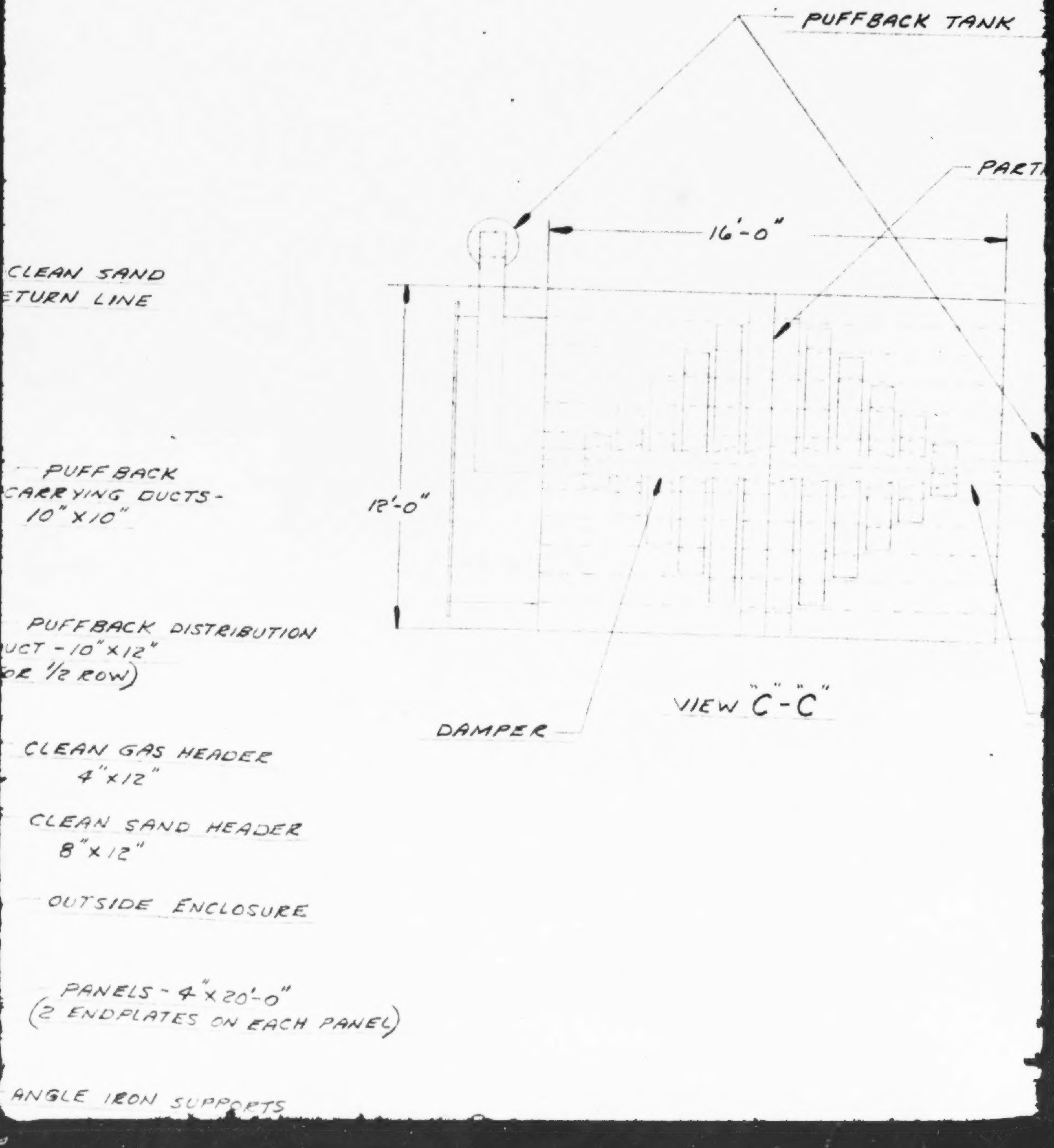
REV. DRAWING NUMBER T440-376

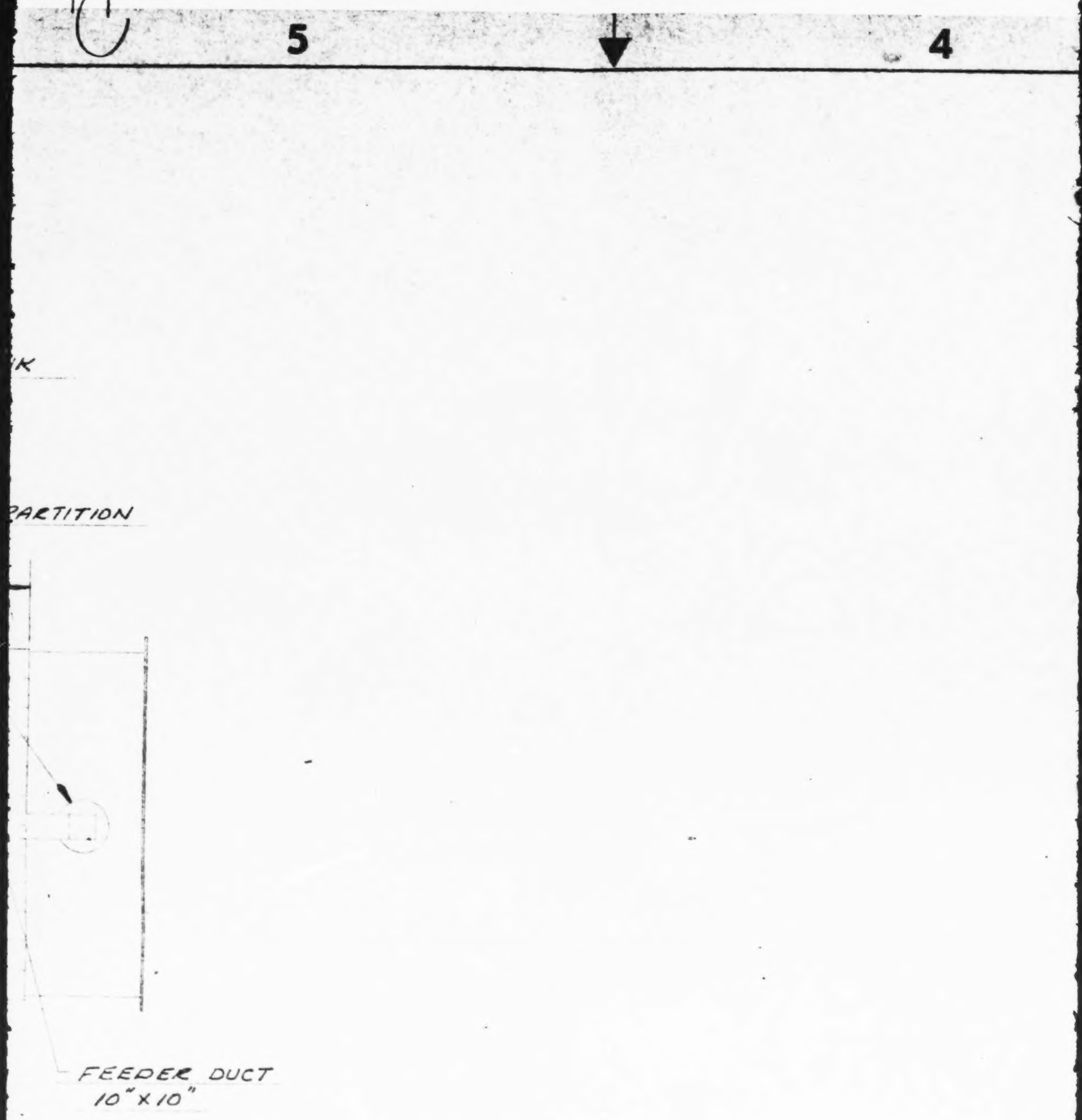
2 DO NOT SCALE DRAWING 1

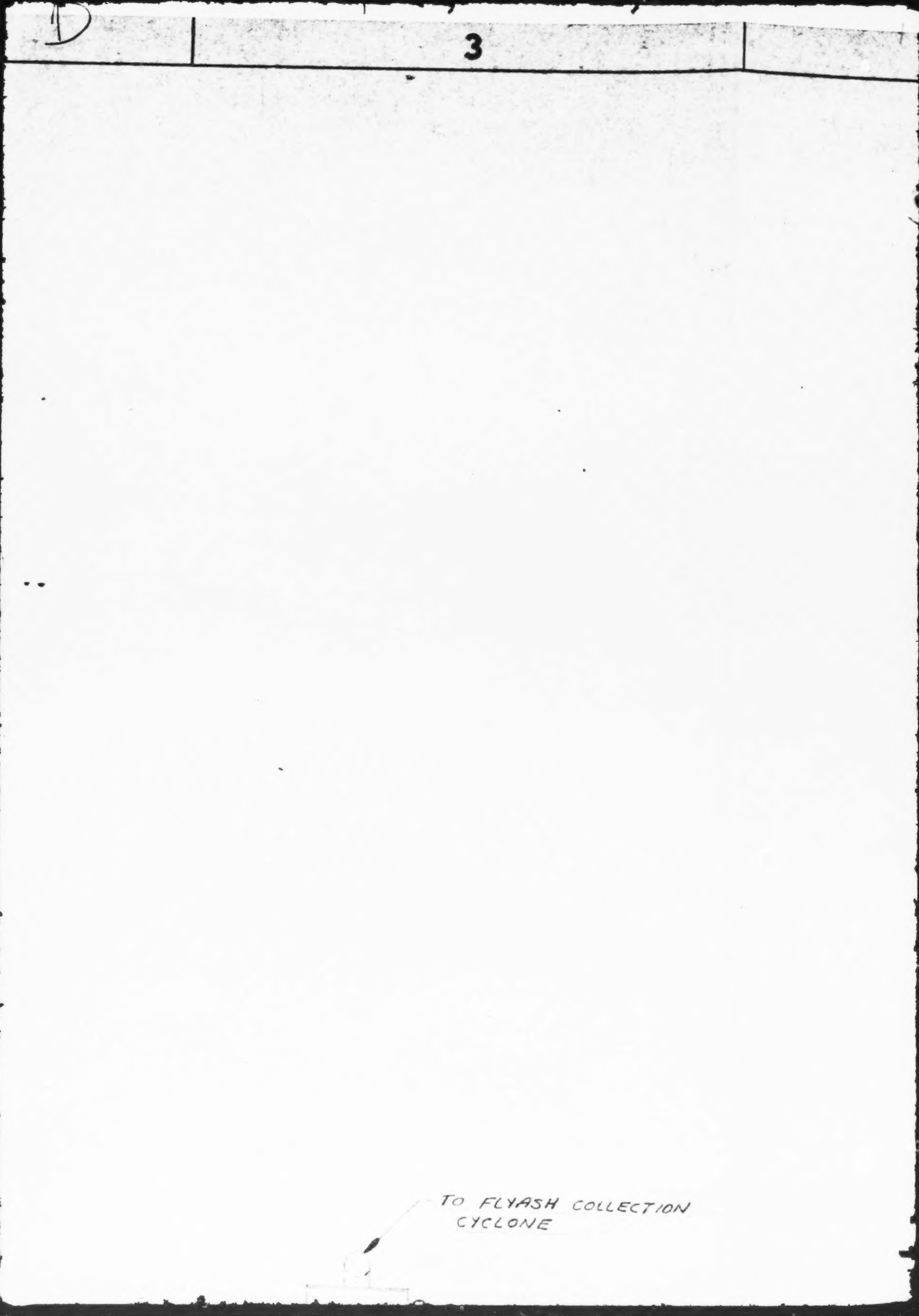

8
7
A
H
G
F
GRANULE FEED BIN
CLEAN S
RETURN L
PUFF
CARRYIN
10" X 1
10"
4"
10"
PUFFE
DUCT - 10
(FOR 1/2 RO
CLEAN
4
CLEA
8" X
6"
OUTS
DAMPERS
PAN
(2 E

7 B

6


PUFFBACK TANK
PART
16'-0"
12'-0"
CLEAN SAND
ETURN LINE
PUFF BACK
CARRYING DUCTS -
10" X 10"
PUFFBACK DISTRIBUTION
UCT - 10" X 12"
OR 1/2 ROW)
DAMPER
VIEW "C"-"C"
CLEAN GAS HEADER
4" X 12"
CLEAN SAND HEADER
8" X 12"
OUTSIDE ENCLOSURE
PANELS - 4" X 20'-0"
(2 ENDPLATES ON EACH PANEL)
ANGLE IRON SUPPORTS


C
5
4
K
PARTITION
FEEDER DUCT
10"×10"


D
3
TO FLYASH COLLECTION CYCLONE

2

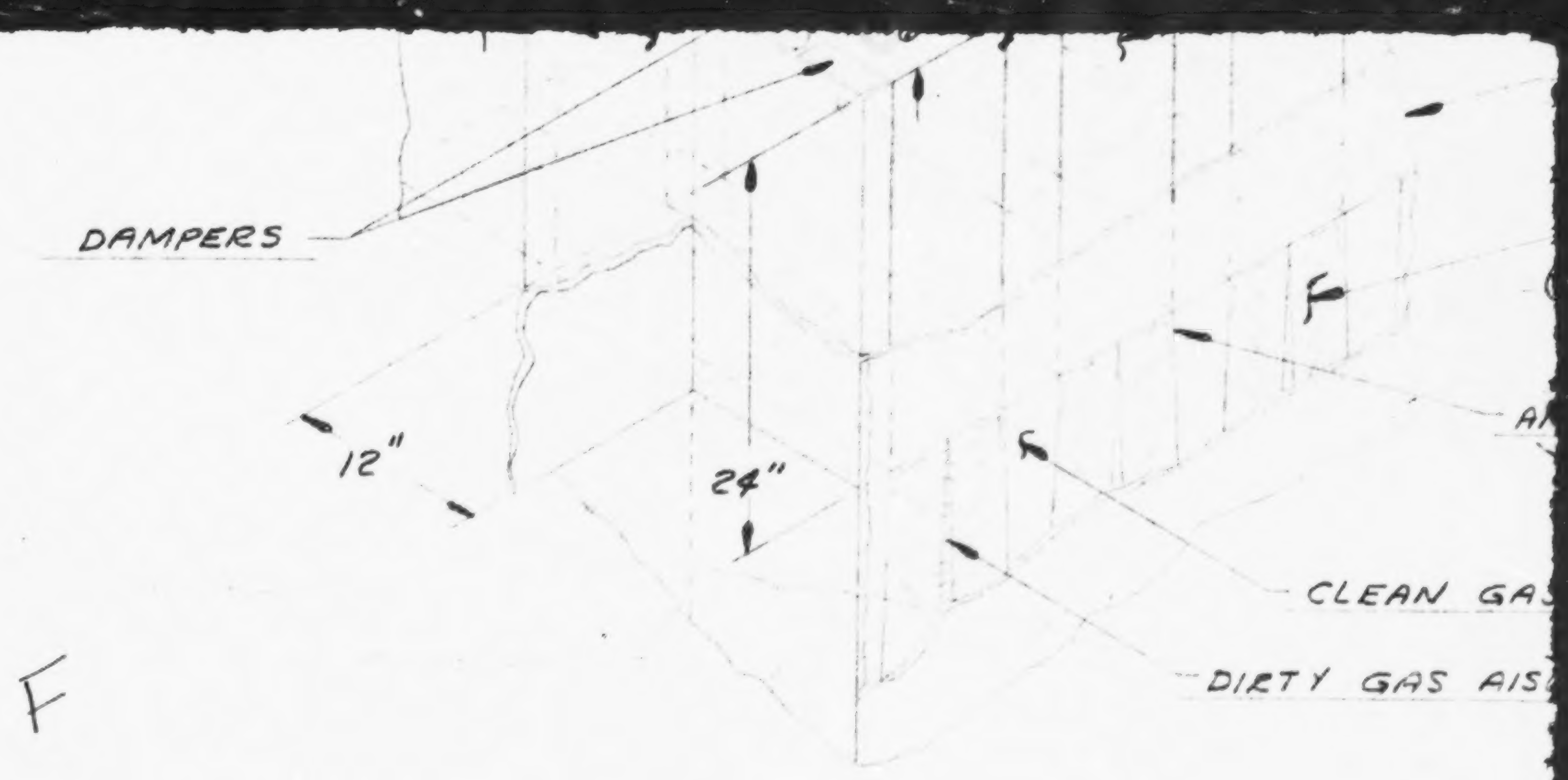
1

ENGR CHG	REVISIONS				
	ZONE	LTR	DESCRIPTION	DATE	APPROVED

H

G

F


DAMPERS
12"
24"
F
F
A
CLEAN GAS
DIRTY GAS AIS

E

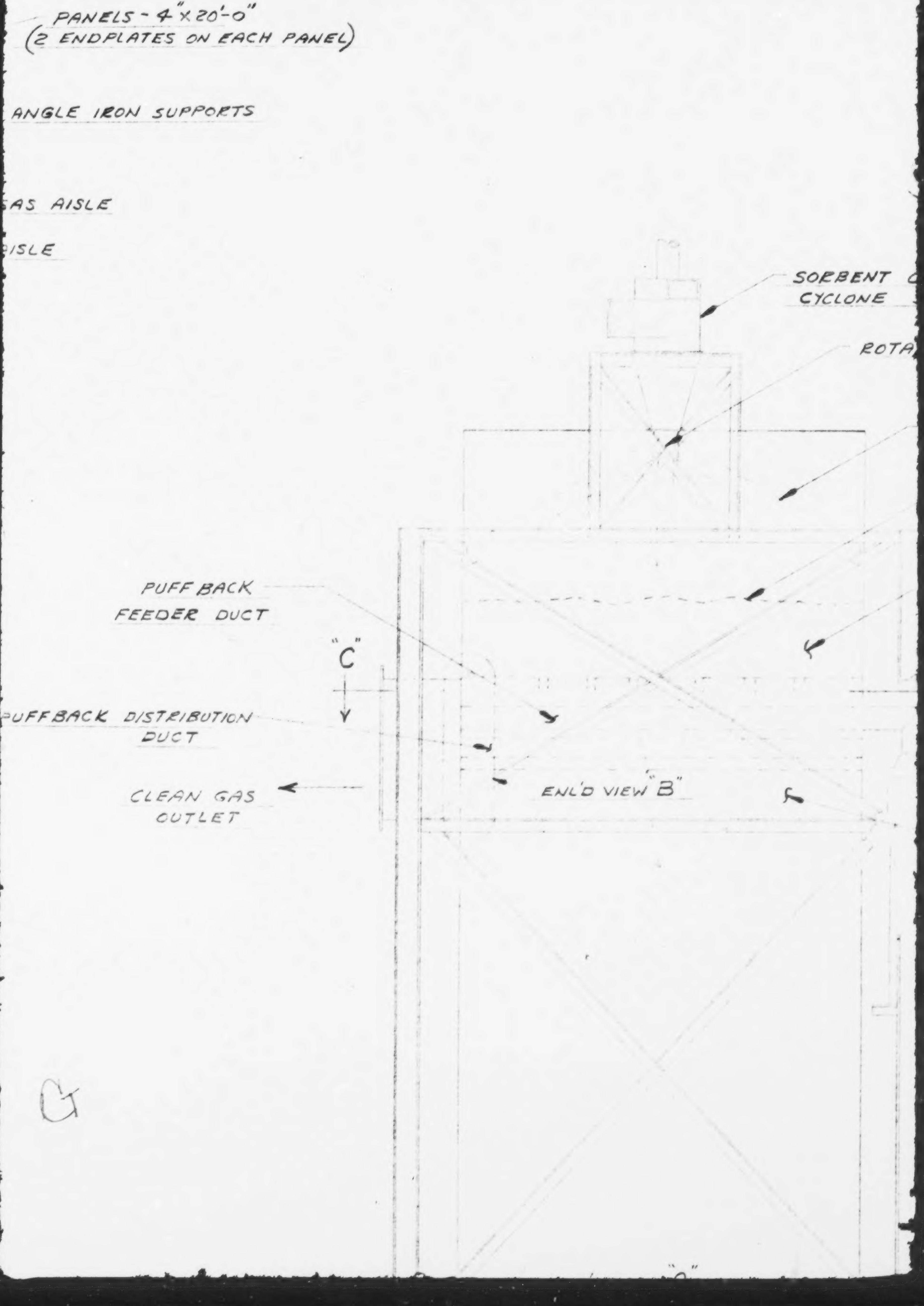
VIEW "B"
SCALE: 1"=1'-0"

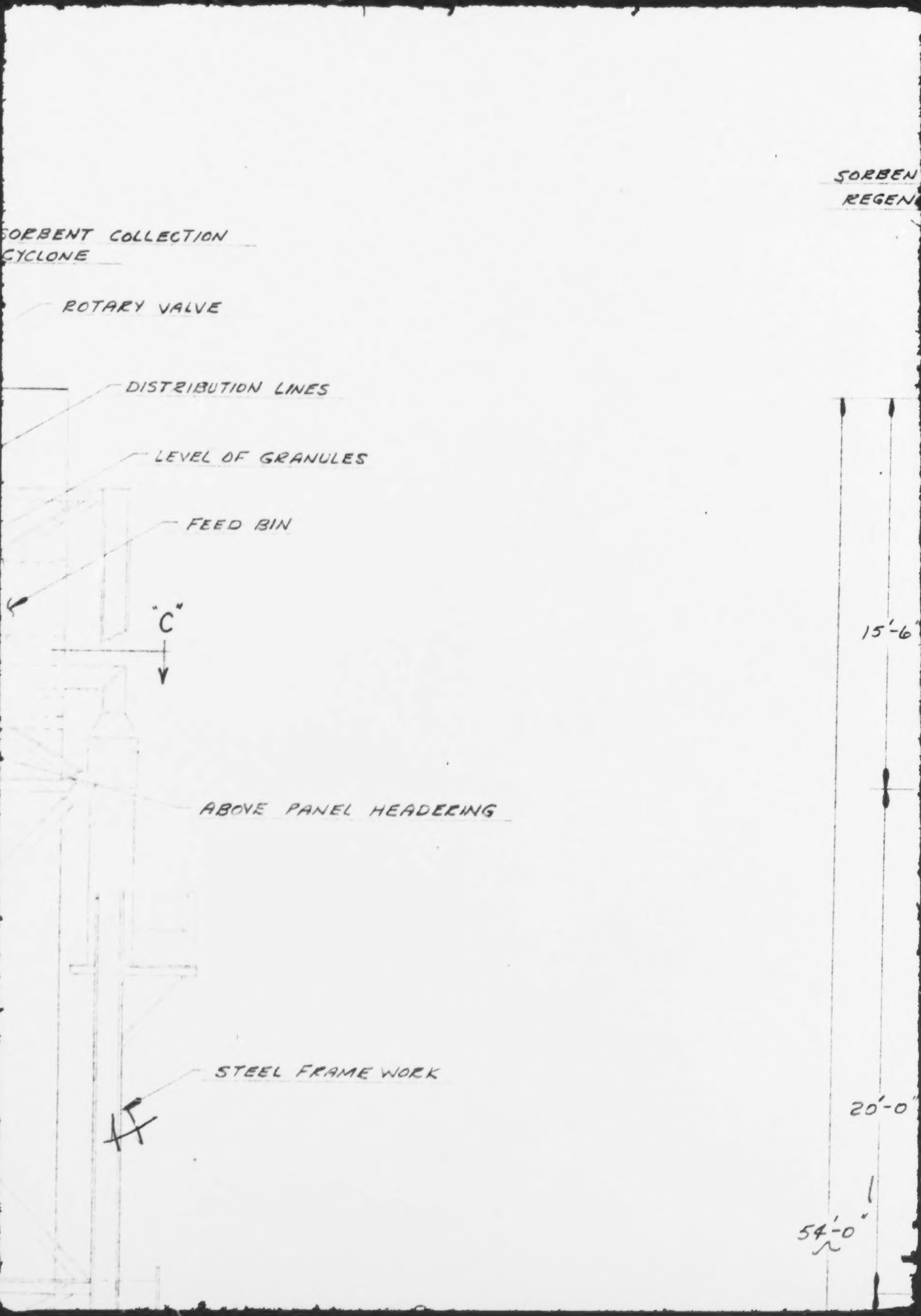
PUF

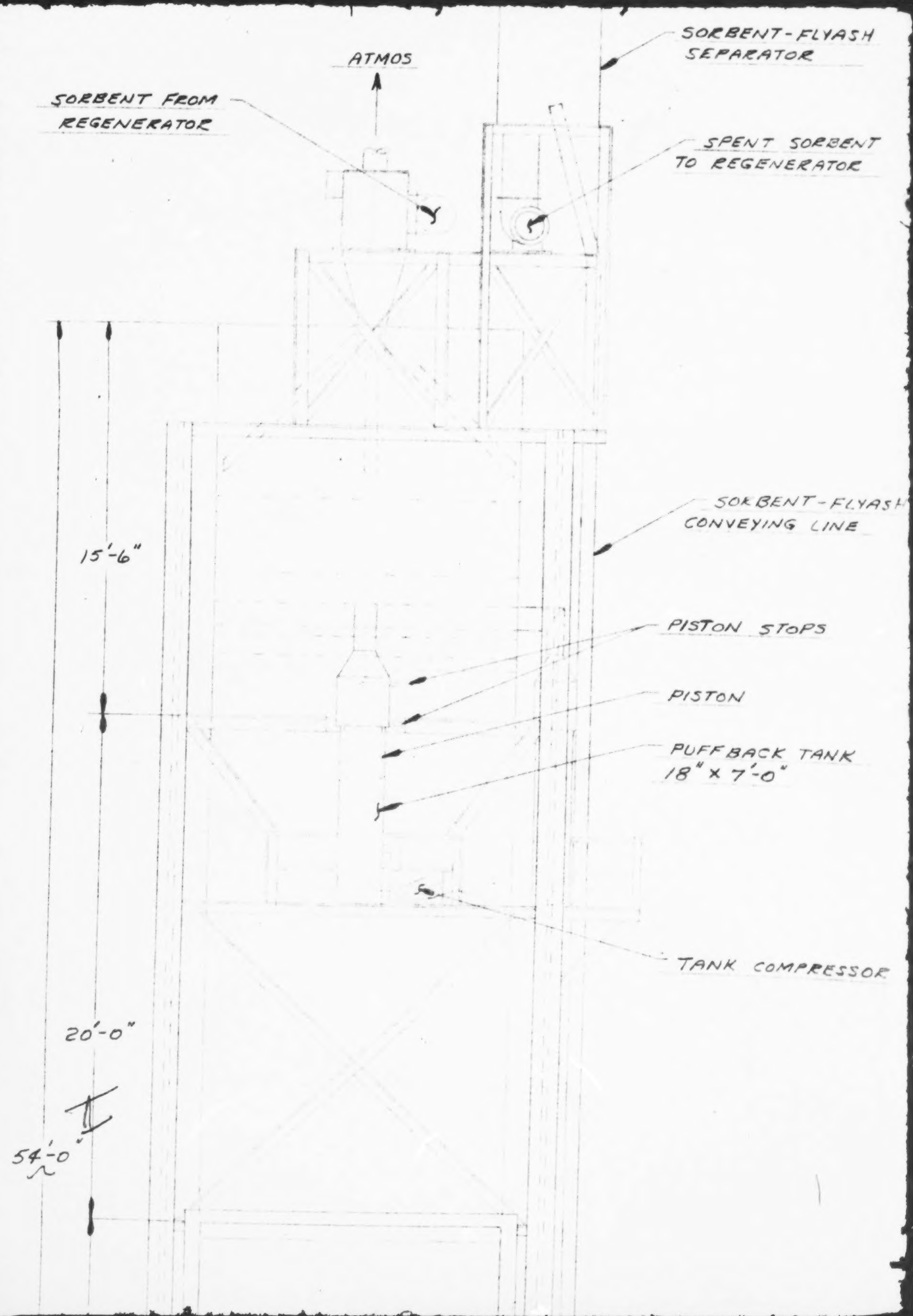
D

CLEAN GAS AISLE
DIRTY GAS AISLE

C


PANELS - 4" x 20'-0"
(2 ENDPLATES ON EACH PANEL)
ANGLE IRON SUPPORTS
AS AISLE
ISLE
SORBENT
CYCLONE
ROTA
PUFF BACK
FEEDER DUCT
"C"
UFF BACK DISTRIBUTION
DUCT
CLEAN GAS
OUTLET
ENL'D VIEW "B"
G


SORBEN
REGEN
SORBENT COLLECTION
CYCLONE
ROTARY VALVE
DISTRIBUTION LINES
LEVEL OF GRANULES
FEED BIN
"C"
15'-6"
ABOVE PANEL HEADERING
STEEL FRAME WORK
20'-0"
54'-0"


SORBENT-FLYASH SEPARATOR
ATMOS
SORBENT FROM REGENERATOR
SPENT SORBENT TO REGENERATOR
SORBENT-FLYASH CONVEYING LINE
15'-6"
PISTON STOPS
PISTON
PUFFBACK TANK 18" x 7'-0"
TANK COMPRESSOR
20'-0"
54'-0"

SPENT SORBENT
TO REGENERATOR

SORBENT-FLYASH
CONVEYING LINE

PISTON STOPS

PISTON

PUFFBACK TANK
18" × 7'-0"

TANK COMPRESSOR

E

REV.

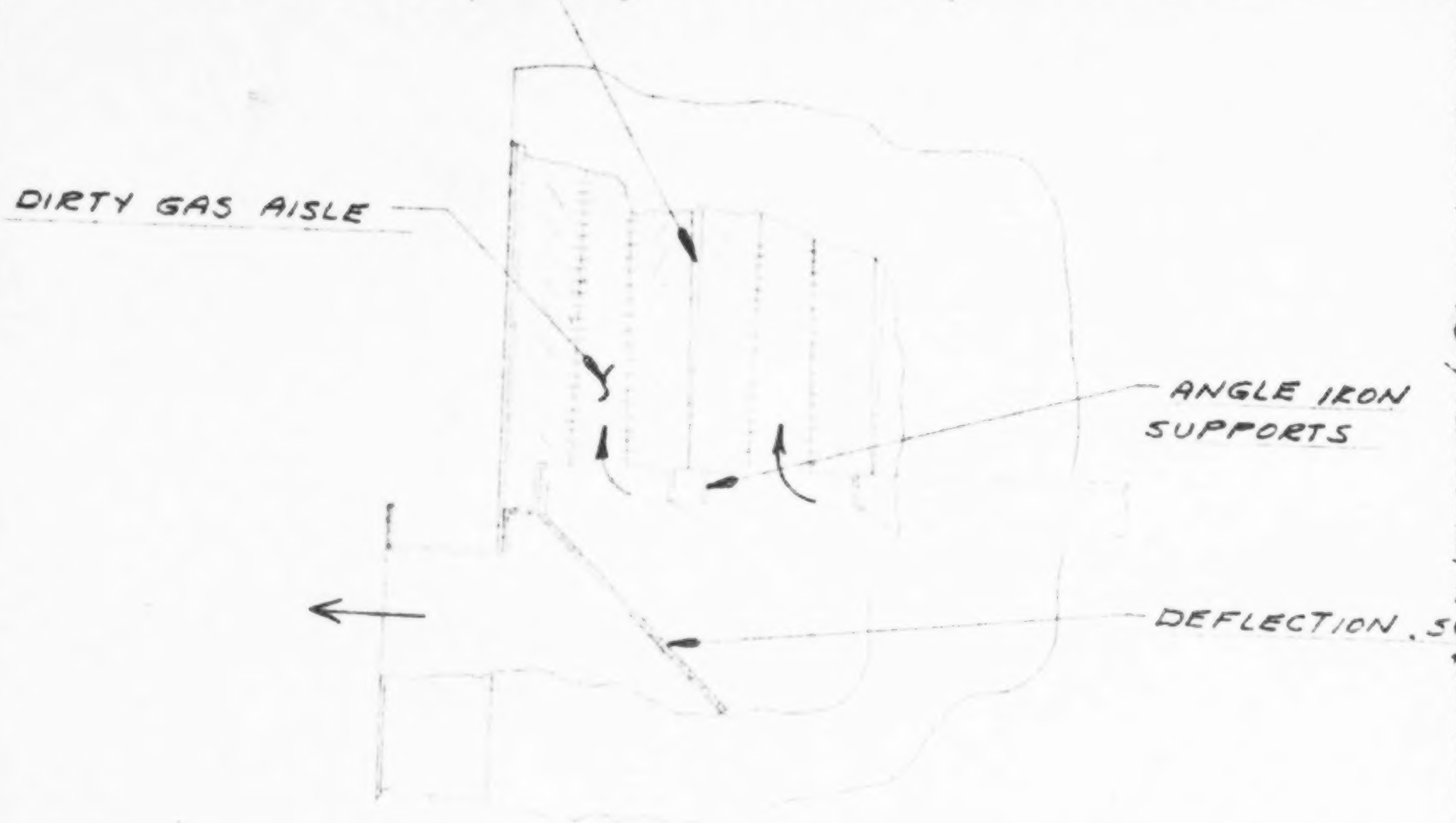
DRAWING NUMBER

D

C

K

C


DIRTY GAS AISLE
ANGLE IRON SUPPORTS
DEFLECTION .S

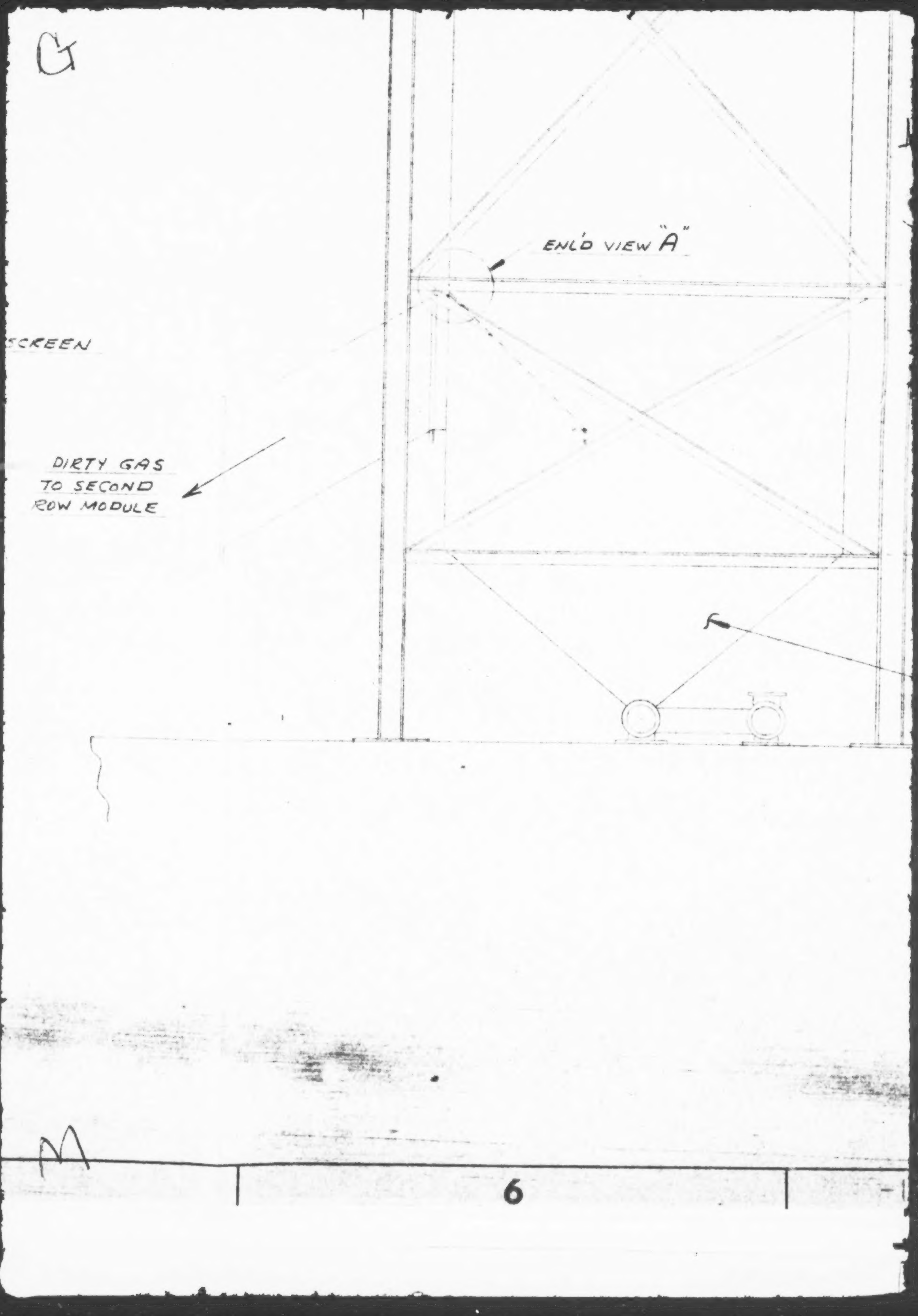
VIEW "A"

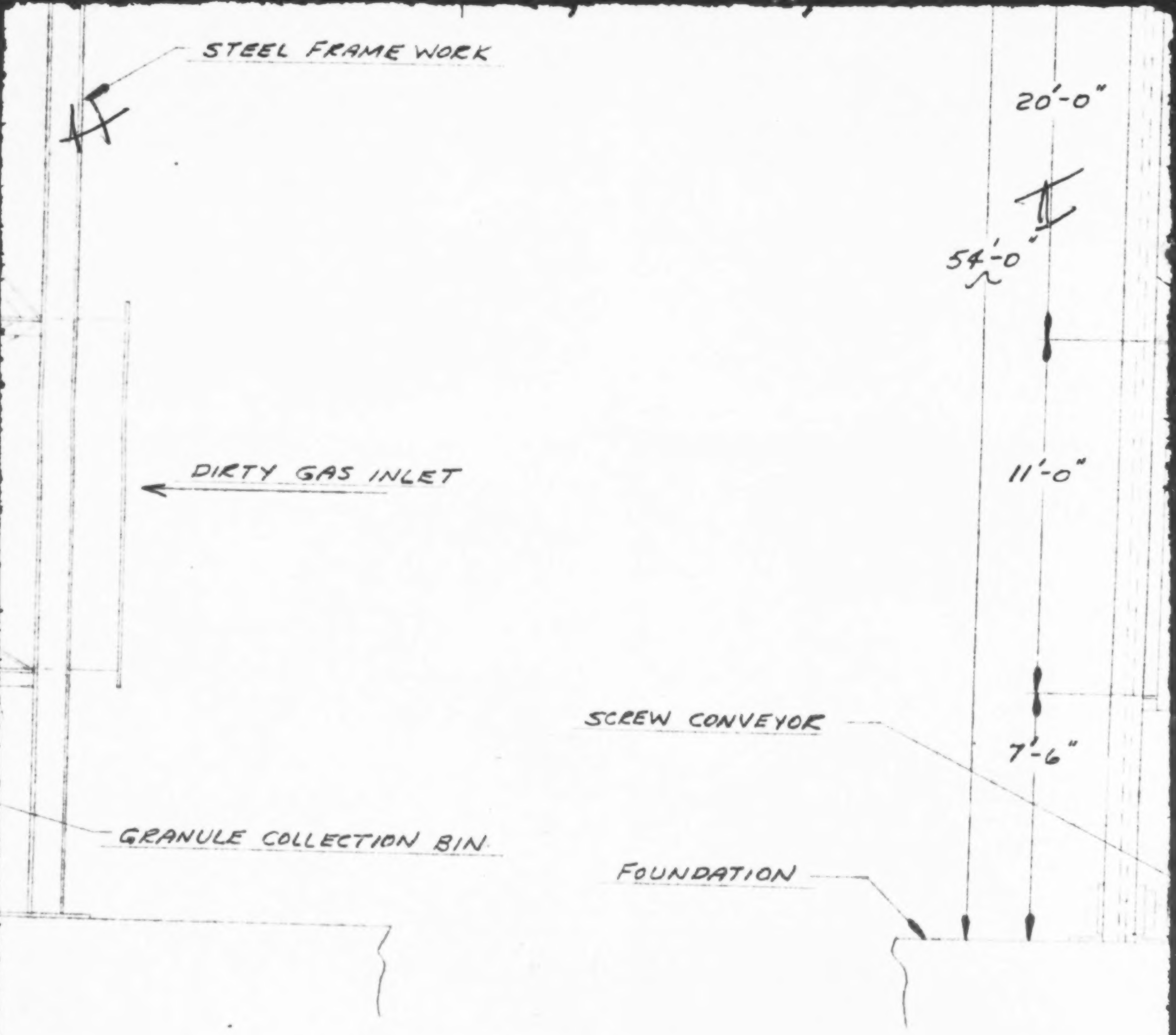
B

A

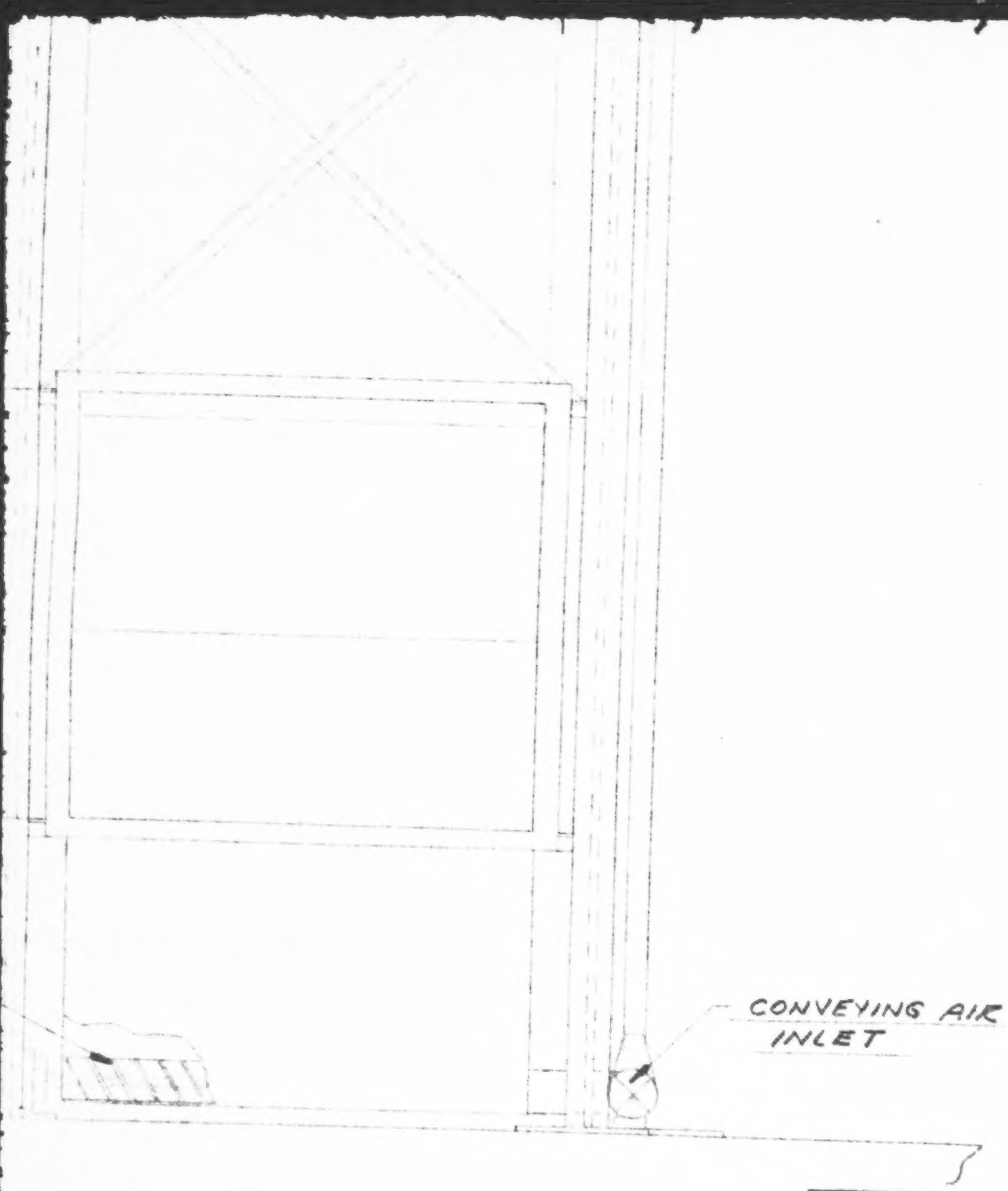
8

7


G
ENL'D VIEW "A"
SCREEN
DIRTY GAS
TO SECOND
ROW MODULE
M
6


STEEL FRAME WORK
20'-0"
54'-0"
DIRTY GAS INLET
11'-0"
SCREW CONVEYOR
7'-6"
GRANULE COLLECTION BIN
FOUNDATION

PART CLASS:
DWG CLASS:
ASG BY:
APPD BY:



QTY REQD	CODE IDENT	PART OR IDENTIFYING NO.
	←ASSY NO.	

DASH NO.	NEXT ASSY	USED ON	QTY PER NHA	END ITEM NUMBER	SERIAL NO.
	APPLICATION			EFFECTIVITY	

-UNLESS OTHERWISE SPEC

TOLERANCE AND DRAWI

INTERPRETATION PER RAD-

.X±.08 .XX±.04 .XX

MACHINED ANGLES±0° 30

SHEET METAL BEND ANGLES

RAD-P10001.

SURFACE ROUGHNESS

C

3

C

1

K

REV.

DRAWING NUMBER T440-377

B

	NOMENCLATURE OR DESCRIPTION	MATERIAL OR NOTE	SPECIFICATION	ZONE	ITEM NO.

LIST OF PARTS

PECIFIED-
WING
AD-D143
XXX±.020
± 30'
SLES±2°
S

CONTR NO.		
RELEASE		
DWG APPD		
CHECKED		
DRAWN	JCC 9-12-68	
DESIGN APPROVAL		
MANAGER		
SECT CH		
DESIGNER		

AVCO CORPORATION FIGURE 16
AVCO SPACE SYSTEMS DIVISION / LOWELL, MASSACHUSETTS

TITLE
SCHEMATIC - INDIVIDUAL
MODULE - G.S.C. PANELS -
115, 200 CFM

SIZE	CODE IDENT NO.	DWG NO.
E	22372	T440 - 377

SCALE 1/4" = 1'-0" WT SHEET

A

2 1